PB85-143253

Economic Evaluation of Advanced Limestone Davy S-H, and Dowa Gypsum-Producing FGD (Flue Gas Desulfurization) Processes

(U.S.) Tennessee Valley Authority
Muscle Shoals, AL

Prepared for

Industrial Environmental Research Lab.
Research Triangle Park, NC

Feb 84

U.S. Department of Commerce
National Technical Information Service
NTIS

PB85-143253

EPA-600/7-84-025
TVA/OP/EDT-82/20
February 1984

Economic Evaluation of Advanced Limestone, Davy S-H, and Dowa Gypsum-Producing FGD Processes

by

R.L. Dotson, J.D. Maxwell,
and T.A. Burnett

TVA, Office of Power
Division of Energy Demonstrations and Technology
Muscle Shoals, Alabama 35660

EPA Interagency Agreement No. 79-D-X0511

EPA Project Officer: Norman Kaplan

Industrial Environmental Research Laboratory
Office of Environmental Engineering and Technology
Research Triangle Park, NC 27711

AIR AND ENERGY ENGINEERING RESEARCH LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
RESEARCH TRIANGLE PARK, NC 27711

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/7-84-025	2.	3. RECIPIENT'S ACCESSION NO. PB85 143253
4. TITLE AND SUBTITLE Economic Evaluation of Advanced Limestone, Davy S-H, and Dowa Gypsum-producing FGD Processes		5. REPORT DATE February 1984
		6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) R. L. Dotson, J. D. Maxwell, and T. A. Burnett		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Tennessee Valley Authority Office of Power Division of Energy Demonstrations and Technology Muscle Shoals, Alabama 35660		10. PROGRAM ELEMENT NO.
		11. CONTRACT/GRANT NO. EPA IAG-79-D-X0511
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Final; 6/79-12/82
		14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES IERL-RTP project officer is Norman Kaplan, Mail Drop 61, 919/541-2556.

16. ABSTRACT The report gives results of economic evaluations of three gypsum-producing flue gas desulfurization processes: advanced limestone (in-loop forced oxidation with adipic acid additive), Davy S-H (lime), and Dowa (aluminum sulfate, limestone). For a 500-MW power unit burning 3.5% sulfur coal and meeting the 1979 New Source Performance Standards, capital investments in 1982 costs are $93 million ($186/kW) for the advanced limestone process, $116 million ($231/kW) for the Davy S-H process, and $121 million ($243/kW) for the Dowa process. First-year annual revenue requirements in 1984 costs for these processes are $26, $33, and $32 million (9.4, 11.9, and 11.7 mills/kWh), respectively. The lower capital investment and annual revenue requirements of the advanced limestone process are due in part to the use of adipic acid, which allows partial scrubbing at 95% removal. The Davy S-H process has slightly higher annual revenue requirements than the Dowa process because lime is used, instead of limestone. Changes in power unit size and coal sulfur content affect the costs of all three processes similarly. The Davy S-H process is more sensitive to raw material costs because lime is used. Landfill waste disposal is a minor cost element in all three processes.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Calcium Carbonates	Pollution Control	13B	07B
Flue Gases	Calcium Oxides	Stationary Sources	21B	
Desulfurization	Aluminum Sulfate	Davy S-H Process	07A,07D	
Coal		Dowa Process	21D	
Gypsum			08G	
Adipic Acid			07C	

18. DISTRIBUTION STATEMENT Release to Public	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 199
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (9-73)

DISCLAIMER

This report was prepared by the Tennessee Valley Authority (TVA) and has been reviewed by the Office of Energy, Minerals, and Industry, U.S. Environmental Protection Agency (EPA), and approved for publication. Neither EPA, TVA, nor any person acting on their behalf:

a. makes any warranty or representation, express or implied, with respect to the use of any information contained in this report, or that the use of any information, apparatus, method, or process disclosed in this report may not infringe privately owned rights; or

b. assumes any liabilities with respect to the use of, or for damages resulting from the use of, any information, apparatus, method, or process disclosed in this report.

This report does not necessarily reflect the views and policies of TVA or EPA.

ABSTRACT

Economic evaluations were made of three gypsum-producing FGD processes: advanced limestone (in-loop forced oxidation with adipic acid additive), Davy S-H (lime), and Dowa (aluminum sulfate, limestone). For a 500-MW power unit burning 3.5% sulfur coal and meeting the 1979 NSPS, capital investments in 1982 costs are 93 M$ (186 $/kW) for the advanced limestone process, 116 M$ (231 $/kW) for the Davy S-H process, and 121 M$ (243 $/kW) for the Dowa process. First-year annual revenue requirements in 1984 costs for these processes are 26, 33, and 32 M$ (9.4, 11.9, and 11.7 mills/kWh), respectively. The lower capital investment and annual revenue requirements of the advanced limestone process is due in part to the use of adipic acid, which allows partial scrubbing at 95% removal. The Davy S-H has slightly higher annual revenue requirements than the Dowa process because lime rather than limestone is used. Changes in power unit size and coal sulfur content affect the costs of all three processes similarly. The Davy S-H process is more sensitive to raw material costs because lime is used. Landfill waste disposal is a minor cost element in all three processes.

CONTENTS

Abstract iii
Figures vii
Tables ix
Abbreviations and Conversion Factors xi

Executive Summary S-1

Introduction 1

Background 3
  Process Evolution 4
  Process Design Variations 4
  Oxidation Chemistry 6
  Additives 7
  Waste Disposal 9
  Adipic Acid-Enhanced Forced-Oxidation Limestone Process 10
  Davy S-H Process 13
  Dowa Process 15

Design and Economic Premises 21
  Design Premises 21
    Coal Premises 21
    Power Plant 21
    Flue Gas Compositions 23
    Environmental Regulations 23
    Particulate Matter 29
    Flue Gas Desulfurization 29
    Solids Disposal 31
  Economic Premises 31
    Schedule and Cost Factors 31
    Capital Cost Estimates 35
    Annual Revenue Requirements 37
  Accuracy of Estimates 40

Systems Estimated 41
  Advanced Limestone Process 41
    Major Processing Areas 41
  Davy S-H Process 54
    Major Processing Areas 54
  Dowa Process 67
    Major Processing Areas 67

Preceding page blank

Results 83
  Capital Investment 83
    Base Case 83
    Case Variations 86
  Annual Revenue Requirements 86
    Base Case 93
    Case Variations 95
  Dowa L/G Ratio Case Variation 99
    Capital Investment 99
    Annual Revenue Requirements 101
  Variations in Economic Factors 101

Conclusions 113

References 115

Appendix A A-1

FIGURES

Number		Page
1	Controlled SO_2 emission requirements for 1979 NSPS	28
2	Landfill plan and construction details	33
3	Advanced limestone process flow diagram	42
4	Advanced limestone process plan and elevation diagram	45
5	Advanced limestone process plot plan	46
6	Davy S-H process flow diagram	55
7	Davy S-H process plan and elevation diagram	58
8	Davy S-H process plot plan	59
9	Dowa process flow diagram	68
10	Dowa process plan and elevation diagram	72
11	Dowa process plot plan	73
12	Power unit size versus total capital investment for new units.	89
13	Power unit size versus unit capital investment for new units.	89
14	Sulfur in coal versus total capital investment	91
15	Sulfur in coal versus unit capital investment	91
16	Power unit size versus levelized annual revenue requirements, new units	98
17	Coal sulfur content versus levelized annual revenue requirements	98
18	Raw material cost versus levelized annual revenue requirements at different power unit sizes	103
19	Raw material cost versus levelized annual revenue requirements at different coal sulfur levels	104
20	Electricity cost versus levelized annual revenue requirements at different power unit sizes	105
21	Variations in electricity cost versus levelized annual revenue requirements at different coal sulfur contents . . .	106
22	Variations in maintenance cost versus levelized annual revenue requirements at different power unit sizes	107
23	Variations in maintenance cost versus levelized annual revenue requirements at different coal sulfur levels	108
24	Variations in labor cost versus levelized annual revenue requirements at different power unit sizes	109
25	Variations in labor cost versus levelized annual revenue requirements at different coal sulfur levels	110
26	Sensitivity to changes in inflation and discount rate	111

TABLES

Number		Page
S-1	Capital Investment Summary	S-6
S-2	Base Case Capital Investment Summary by Processing Areas	S-7
S-3	Annual Revenue Requirement Summary	S-9
S-4	Major Cost Components for Base Case Annual Revenue Requirements	S-10
1	Coal Compositions	22
2	Power Unit Operating Time and Heat Rate	23
3	Flue Gas Composition for 5.0% Sulfur Eastern Bituminous Coal	24
4	Flue Gas Composition for 3.5% Sulfur Eastern Bituminous Coal	25
5	Flue Gas Composition for 2.0% Sulfur Eastern Bituminous Coal	26
6	1979 Revised NSPS Emission Standards	27
7	Premise Coal Emission Standards	27
8	Raw Material Characteristics	32
9	Cost Indexes and Projections	35
10	Cost Factors	34
11	Indirect Capital Cost Factors	36
12	Contingency and Allowance for Startup and Modifications Cost Factors	36
13	Advanced Limestone Process Material Balance	43
14	Advanced Limestone Process Equipment List	47
15	Davy S-H Process Material Balance	56
16	Davy S-H Process Equipment List	60
17	Dowa Process Material Balance	69
18	Dowa Process Equipment List	74
19	Capital Investment Summary	84
20	Capital Investment by Processing Areas for New 500-MW Power Units with 3.5% Sulfur Coal	85
21	Capital Investment by Processing Area for New 200-, 500-, and 1,000-MW Power Units with 3.5% Sulfur Coal	87
22	Capital Investment by Processing Area for Existing 200-, 500-, and 1,000-MW Power Units with 3.5% Sulfur Coal	88
23	Capital Investment by Processing Area for New 500-MW Power Units with 2.0%, 3.5%, and 5.0% Sulfur Coal	90
24	Annual Revenue Requirement Summary	92
25	Major Cost Components for Base Case Annual Revenue Requirements	94

Preceding page blank

TABLES (Continued)

Number		Page
26	Annual Revenue Requirement Summary for New 200-, 500-, and 1,000-MW Power Units with 3.5% Sulfur Coal	96
27	Annual Revenue Requirement Summary for Existing 200-, 500-, and 1,000-MW Power Units with 3.5% Sulfur Coal	97
28	Annual Revenue Requirement Summary for New 500-MW Power Units with 2.0%, 3.5%, and 5.0% Sulfur Coal	100
29	Annual Revenue Requirement Cost Element Variations	101
30	Effect of Capital Charges Variations on Base Case Capital Investment and Levelized Annual Revenue Requirements	112

ABBREVIATIONS AND CONVERSION FACTORS

ABBREVIATIONS

aft3/min	actual cubic feet per minute	lb	pound
Btu	British thermal unit	L/G	liquid-to-gas ratio in gallons per thousand actual cubic feet of gas at outlet conditions
$^{\circ}$F	degrees Fahrenheit		
dia	diameter		
FGD	flue gas desulfurization	M	million
ft	feet	mi	mile
ft^2	square feet	mo	month
ft^3	cubic feet	MW	megawatt
gal	gallon	ppm	parts per million
gpm	gallons per minute	psig	pounds per square inch (gauge)
gr	grain		
hp	horsepower	rpm	revolutions per minute
hr	hour	sec	second
in.	inch	sft3/min	standard cubic feet per minute (60°F)
k	thousand		
kW	kilowatt	SS	stainless steel
kWh	kilowatthour	yr	year

CONVERSION FACTORS

EPA policy is to express all measurements in Agency documents in metric units. Values in this report are given in British units for the convenience of engineers and other scientists accustomed to using the British systems. The following conversion factors may be used to provide metric equivalents.

To convert British		Multiply by	To obtain Metric	
ac	acre	0.405	hectare	ha
Btu	British thermal unit	0.252	kilocalories	kcal
°F	degrees Fahrenheit minus 32	0.5556	degrees Celsius	°C
ft	feet	30.48	centimeters	cm
ft^2	square feet	0.0929	square meters	m^2
ft^3	cubic feet	0.02832	cubic meters	m^3
ft/min	feet per minute	0.508	centimeters per second	cm/s
ft^3/min	cubic feet per minute	0.000472	cubic meters per second	m^3/s
gal	gallons (U.S.)	3.785	liters	L
gpm	gallons per minute	0.06308	liters per second	L/s
gr	grains	0.0648	grams	g
gr/ft^3	grains per cubic foot	2.288	grams per cubic meter	g/m^3
hp	horsepower	0.746	kilowatts	kW
in.	inches	2.54	centimeters	cm
lb	pounds	0.4536	kilograms	kg
lb/ft^3	pounds per cubic foot	16.02	kilograms per cubic meter	kg/m^3
lb/hr	pounds per hour	0.126	grams per second	g/s
psi	pounds per square inch	6895	pascals (newton per square meter)	Pa (N/m^2)
mi	miles	1609	meters	m
rpm	revolutions per minute	0.1047	radians per second	rad/s
sft^3/min	standard cubic feet per minute (60°F)	1.6077	normal cubic meters per hour (0°C)	m^3/hr (0°C)
ton	tons (short)	0.9072	metric tons	tonne
ton/hr	tons per hour	0.252	kilograms per second	kg/s

ECONOMIC EVALUATION OF ADVANCED LIMESTONE, DAVY S-H, AND DOWA GYPSUM-PRODUCING FGD PROCESSES

EXECUTIVE SUMMARY

INTRODUCTION

Since 1967, the Tennessee Valley Authority (TVA) has conducted economic evaluations of flue gas desulfurization (FGD) processes and related technology of electric utility power plants. Many of these evaluations have been sponsored by the U.S. Environmental Protection Agency (EPA) as part of its FGD research and development program. The major objective of each evaluation is to prepare economic analyses which are representative of current U.S. utility power plant conditions on a consistent basis that allows equitable process comparisons.

This study for EPA is the fifth in a series begun in 1977 to evaluate the economics of advanced FGD processes that have promise of commercial development. Three gypsum-producing FGD processes with landfill disposal are evaluated in this study: a limestone-scrubbing process with adipic acid addition and forced oxidation (the advanced limestone process), the Davy Saarberg-Holter (S-H) process, and the Dowa process.

Although the technology is not new, production of gypsum rather than calcium sulfite sludge in limestone scrubbing is an emerging FGD technology which has only recently attracted wide commercial interest in the United States. In Japan, many limestone and lime FGD systems produce gypsum by the forced-oxidation method since there are few natural gypsum deposits and the byproduct FGD gypsum can be sold to the wallboard and cement industries. In the United States, the major emphasis has not been on producing gypsum for sale but to solve the waste problems associated with disposal of calcium sulfite sludge. Calcium sulfite sludge forms an unstable, semiliquid material when disposed of in a pond. It is difficult to dewater and must be stabilized to be disposed of by landfill. Gypsum slurries are easier to dewater and can be disposed of in a landfill without stabilization. A gypsum slurry can be dewatered to 80% to 85% solids with a conventional vacuum filter. The filter cake resembles a moist soil that is easily handled by conventional solids-handling equipment. Although some byproduct FGD gypsum can probably be sold to U.S. wallboard and cement manufacturers, this evaluation assumes that all waste gypsum is landfilled.

PROCESS DESCRIPTION

The advanced limestone process is similar to the basic limestone slurry process except for two modifications: (1) adipic acid is added to the scrubbing liquid to enhance SO_2 removal and increase limestone utilization and (2) forced oxidation by air sparging is incorporated to produce gypsum. The design and operating conditions are based in part on data from EPA-sponsored tests at the Shawnee test facility. The process uses a countercurrent spray tower with a presaturator and a chevron-type mist eliminator. The presaturator cools the entering gas from 300°F to 124°F. The absorbent liquid is an 8% solids slurry of finely ground (90% minus 325 mesh) limestone prepared onsite by crushing and ball milling. A stoichiometry of 1.07 mols $CaCO_3$/mol (SO_2 + 2HCl) absorbed is used. The liquid-to-gas (L/G) ratio is 4 gal/kaft^3 for the presaturator and 80 gal/kaft^3 for the absorber. An adipic acid concentration of 1,500 ppm is maintained in the absorption loop by metering adipic acid into the makeup limestone slurry tank. Air at a stoichiometry of 2.5 lb atoms O/lb mol SO_2 absorbed is injected into the absorber effluent tank to oxidize the sulfite. A bleedstream is pumped to a thickener and rotary vacuum filter for dewatering. The dewatered gypsum is trucked to a landfill one mile from the FGD facility.

The Davy S-H process is a lime-based process using a clear alkaline absorbent liquid buffered with formic acid. The SO_2 is removed in a Rotopart absorber (a patented modular design cocurrent spray column with an integral gas-liquid separator). Each module consists of two absorption tubes containing liquid-shedding rings (venturi throats) connected to a single gas-liquid separator (Rotopart) section. The buffering action of formic acid acts to slow and control the pH drop from 4.5 to 5.0, assuring formation of bisulfite. The concentrations of both the formic acid and the ferric chloride catalyst and the absorber L/G ratio are proprietary information. The lime stoichiometric ratio is about 1.02 mols CaO/mol (SO_2 + 2HCl) absorbed. The absorber effluent is oxidized by air sparging in an oxidizer vessel at a stoichiometry of 5 lb atoms O/lb mol SO_2 absorbed. The gypsum slurry overflows from the oxidizer to the thickener through a mixing trough. Lime slurry and makeup formic acid are added to this trough. The thickened slurry is dewatered in a rotary vacuum filter. The dewatered gypsum is trucked one mile to a landfill.

The Dowa process is a double-alkali process that uses a basic aluminum sulfate [$Al_2(SO_4)_3$] solution to absorb SO_2 and limestone to precipitate gypsum and regenerate the scrubbing solution. A packed-bed countercurrent absorber with a fresh water wash on the wall at the gas-liquid interface is used for SO_2 removal. The absorber is packed with Ecodyne polygrid packing. A L/G ratio of 45 gal/kaft^3 is used. The absorber effluent is oxidized by air sparging in an oxidizer vessel in the absorber liquid loop at an air stoichiometry of 4 lb atoms O/lb mol SO_2 absorbed. A chloride purge is provided to prevent the excessive buildup of chlorides in the system. A bleedstream from the oxidation tank is neutralized with limestone to regenerate the basic $Al_2(SO_4)_3$ solution and precipitate gypsum. A stoichiometry of 1.03 mols $CaCO_3$/mol of (SO_2 + 2HCl) absorbed is used. The gypsum slurry is dewatered in a thickener and a rotary vacuum filter, and the gypsum is trucked one mile to a landfill for disposal.

DESIGN AND ECONOMIC PREMISES

The economic evaluations are based on a conceptual design developed from the design premises and engineering data such as flow diagrams, material balances, and equipment costs. The Davy S-H and Dowa processes are based on vendor-supplied data; the limestone process is a generic design. A base case, new 500-MW power unit burning a 3.5% sulfur, 16% ash (both are dry basis) bituminous coal, and complying with the 1979 new source performance standards (NSPS) is used as the primary basis of comparison.

Design Premises

The coals used are three eastern bituminous coals containing 5.0%, 3.5%, and 2.0% sulfur (dry basis). The coals have a heating value of 11,700 Btu/lb as fired. The compositions of the coals are shown below. The 3.5% sulfur coal is the base case coal.

Coal	Sulfur, %		Ash, %	Moisture, %	Heat content, Btu/lb
	Dry basis	As-fired basis	As-fired basis		
Eastern bituminous	5.0	4.80	15.10	4.0	11,700
Eastern bituminous	3.5	3.36	15.14	4.0	11,700
Eastern bituminous	2.0	1.92	15.08	4.0	11,700

The power plant site is assumed to be in the north-central region of the United States. The base case power unit is a new, single 500-MW, balanced-draft, horizontally fired, dry-bottom boiler burning pulverized coal. Power unit size variations consist of similar 200- and 1,000-MW units. The power units have a 30-year life and operate at full load for 5,500 hr/yr for a total lifetime operation of 165,000 hours. Existing power units with 20 years of remaining life at 5,500 hr/yr of full-load operation are also evaluated. This operating schedule is equivalent to a total remaining lifetime operation of 110,000 hours.

Flue gas compositions are based on a total air rate equivalent to 139% of the stoichiometric requirement, including 19% air leakage. It is assumed that 80% of the ash in the coal is emitted as fly ash and 92% of the sulfur is emitted as SO_x. Three percent of the sulfur emitted as SO_x is SO_3 and the remainder is SO_2. Cold-side electrostatic precipitators (ESP) sized to meet the 0.03 lb/MBtu 1979 NSPS are assumed for particulate matter control. The costs for ash collection and disposal are not included in the economic evaluation.

SO_2 emissions from new coal-fired utility plants are regulated by the revised 1979 NSPS. SO_2 emissions from existing units are based on a representative state implementation plan (SIP) emission standard of 0.6 lb SO_2/MBtu. The controlled outlet SO_2 emission and SO_2 removal efficiencies for new plants for each of the coals are shown below.

Coal	Equivalent SO_2 content of coal, lb SO_2/MBtu	Overall equivalent SO_2 reduction required, %	Equivalent SO_2 removal required in FGD system, %	Controlled outlet emission, lb SO_2/MBtu
Eastern bit., 5.0% S	8.21	90.0	89.1	0.82
Eastern bit., 3.5% S	5.74	89.6	88.7	0.60
Eastern bit., 2.0% S	3.28	81.7	80.1	0.60

The FGD systems are coupled to the power unit by a plenum into which the power plant induced-draft (ID) fans discharge. Separate ID fans are provided for the FGD system. The FGD costs include ductwork and associated equipment between the power plant ID fans and the stack plenum. Costs for the stack plenum and the stack are excluded. Absorption trains are sized for a maximum of 125 MW of flue gas (513,000 sft^3/min). Spare trains are provided for a minimum of 25% redundancy and provision for emergency bypass of 50% of the scrubbed gas is included. For all of the advanced limestone cases, partial scrubbing of the flue gas at a SO_2 removal efficiency of 95% SO_2 is used. Partial scrubbing of 89% of the flue gas at 90% SO_2 removal is used for the 2.0% sulfur coal cases for the Davy S-H and the Dowa processes. This corresponds to the quantity of gas bypassed shown below. Indirect steam reheat to 175°F is provided as necessary. The dewatered gypsum cake is disposed of in a landfill located one mile from the FGD facilities. The gypsum disposal area includes loading, trucking, spreading and compaction operations, and general landfill maintenance and reclamation.

Coal	Flue gas bypass, %		
	Advanced limestone	Davy S-H	Dowa
Eastern bit., 5.0% S	6.2	-	-
Eastern bit., 3.5% S	6.7	-	-
Eastern bit., 2.0% S	15.7	11.0	11.0

Economic Premises

A 3-year construction period, from early 1981 to late 1983, is used for both new and existing units. Mid-1982 costs are used for capital investment. Mid-1984 costs are used for annual revenue requirements.

Capital investment consists of direct investment, indirect investment, and other capital investment. Direct investment consists of the installed costs of all process equipment, including provisions for services, utilities, and miscellaneous; and the costs associated with the landfill. Indirect capital investment consists of fees for engineering design and supervision, architect and engineering contractor, construction expense, contractor fees, and contingency. An allowance for startup and modifications and interest during construction is included. No royalty charges are included for the advanced limestone and Dowa processes. Royalty charges for the Davy S-H process were obtained from the process vendor. Process and waste disposal area land is charged at $5,000 per acre.

For existing plant cases, a capital investment retrofit factor of 1.3 is assigned to cover the additional investment required for fitting the equipment into the available space. Each of the area investments (e.g., materials handling, etc.) is multiplied by the retrofit factor. The remaining investment costs are calculated in the same manner as for new units.

Annual revenue requirements consist of various direct and indirect operating and maintenance costs and capital charges. Both first-year and levelized annual revenue requirements are reported. Levelized capital charges are used in both first-year and levelized annual revenue requirements. Levelized operating and maintenance costs are determined by multiplying the first-year operating and maintenance costs by a 30-year levelizing factor of 1.886. This represents an annual inflation rate of 6% and a discount rate of 10%.

SYSTEMS ESTIMATED

All three processes are divided into materials handling, feed preparation, gas handling, SO_2 absorption, reheat, oxidation, solids separation, and disposal areas. In the Dowa process, two additional areas, neutralization and chloride purge, are included. There are no equivalent areas in the other processes. The materials handling and feed preparation areas consist of equipment to receive, store, and prepare absorbent slurry. Limestone is crushed and wet ball milled. Lime is slaked and slurried. For processes that use additives, supplemental equipment is included in the preparation area for addition. The gas-handling area consists of a feed plenum that distributes the flue gas to the individual absorber trains, ductwork, and one ID fan per absorber train. Two emergency bypass ducts are included to handle 50% of the scrubbed gas and to serve as bypass ducts if some of the flue gas is normally bypassed. The absorption area consists of the absorbers and the absorbent liquid recirculation system. All FGD systems are designed for a flue gas temperature of 175°F at the entrance to the stack. Reheat is provided by indirect, inline steam reheat with provisions to account for compression by the ID fans and heating by bypassed flue gas. The oxidation area consists of air compressors, the oxidation tank, a sparging ring, and an agitator. The neutralization and chloride purge areas in the Dowa process include tanks, agitators, pumps, and limestone addition equipment to precipitate gypsum and aluminum hydroxide. The solids separation area consists of thickeners and rotary vacuum filters to generate an 85% solids gypsum waste. Trucks are used to transport the filter cake to the landfill.

RESULTS

The economics developed consist of capital investments and first-year and levelized annual revenue requirements for the three processes, based on the design and economic premises and the specific process designs.

Capital Investment

A summary of the capital investments for each process is shown in Table S-1. The advanced limestone process has the lowest capital investment, followed by the Davy S-H process (25% higher in the base case), and the Dowa process (31% higher in the base case).

TABLE S-1. CAPITAL INVESTMENT SUMMARY

	Advanced limestone		Davy S-H		Dowa	
	M$	$/kW	M$	$/kW	M$	$/kW
New units[a]						
200 MW, 3.5% S	54.4	272	63.0	315	66.4	332
500 MW, 2.0% S	79.1	158	99.8	200	102.6	205
500 MW, 3.5% S	92.8	186	115.5	231	121.3	243
500 MW, 5.0% S	97.9	196	121.4	243	130.3	261
1,000 MW, 3.5% S	155.5	156	206.7	207	216.1	216
Existing units[b]						
200 MW, 3.5% S	59.8	299	80.6	403	85.4	427
500 MW, 3.5% S	118.5	237	148.1	296	156.2	312
1,000 MW, 3.5% S	211.7	212	268.1	268	281.3	281

a. New FGD facilities are constructed simultaneously with the power plant and have a 30-year remaining life.

b. Existing FGD facilities are retrofit installations on existing power plant facilities and have a 20-year remaining life.

Direct capital investments by process area for the base case processes are shown in Table S-2. The Davy S-H process has the highest gas handling area cost (due to the limited but costly use of 316 stainless steel ducting) and oxidation area cost (due to a unique sparger constructed with 316 stainless steel). The Davy S-H process has the lowest feed preparation area cost (due to the use of lime) and SO_2 absorption area cost (due to the modular Rotopart). However, the high cost areas of the Davy S-H process are not offset by the lower cost areas, and, therefore, result in the Davy S-H process having the second highest capital investment. The Dowa process has only two major areas (SO_2 absorption and solids separation) that are lower in cost than the advanced limestone process, and they are only marginally lower in cost. The higher costs in the other processing areas result in the Dowa process having the highest capital investment for the base case.

TABLE S-2. BASE CASE CAPITAL INVESTMENT SUMMARY

BY PROCESSING AREAS[a]

	M$		
	Advanced limestone	Davy S-H	Dowa
Materials handling	2.6	3.5	2.9
Feed preparation	4.3	1.2	5.3
Gas handling	10.5	18.6	14.3
SO_2 absorption	17.7	14.7	17.6
Reheat	3.2	4.2	4.0
Oxidation	2.3	5.0	4.7
Neutralization	-	-	0.6
Chloride purge	-	-	0.8
Solids separation	3.5	2.6	2.7
Total process capital	44.1	49.8	52.9
Total direct investment	51.0	57.0	60.3
Total capital investment	92.8	115.5	121.3

a. 500-MW, new power unit, 3.5% sulfur coal, 1982 dollars.

Case Variations--

As shown in Table S-1, the capital investment ranking remains the same for all case variations as that of the base cases. The capital investment approximately doubles as the power unit size increases from 200 to 500 MW and triples as the size increases from 200 to 1,000 MW. For the existing units, the cost relationships are essentially the same as for new units. The capital investment increases 15% to 18% as the coal sulfur increases from 2.0% to 3.5%. The increase is 22% to 27% as the coal sulfur increases from 2.0% to 5.0%.

The 200-MW units have a 50% spare capacity because only two operating absorption trains (arbitrarily used as the minimum) are required, while the 500- and 1,000-MW units have only 25% spare capacity. The additional capital required for the 200-MW unit increases the cost in terms of $/kW, resulting in an exaggerated economy of scale as the power unit size increases from 200 to 500 MW. Partial scrubbing is used in all processes for the 2% sulfur coal. In the advanced limestone cases, the SO_2 absorption and gas-handling equipment are sized to handle 84.3% of the total gas and the remaining 15.7% is bypassed. In the Davy S-H and Dowa processes, the SO_2 absorption equipment and gas-handling equipment are sized to handle 89% of the total gas and the remaining 11% is bypassed. The reduction in equipment size and materials because of the bypass further reduces the capital investment for the 2.0% sulfur coal cases compared with the 3.5% and 5.0% sulfur cases for the Davy S-H and the Dowa processes. For the advanced limestone process where partial scrubbing is used in all cases, the corresponding reductions are not as significant.

Annual Revenue Requirements

As shown in Table S-3 the advanced limestone process also has the lowest annual revenue requirements, but there is little difference between the Davy S-H and Dowa processes. Although the Dowa process has a higher capital investment, it has marginally lower annual revenue requirements because it uses limestone (at 8.50 $/ton) whereas, the Davy S-H uses lime at 75 $/ton. The Davy S-H process lime cost is 4.0 and 3.3 times higher than the limestone costs of the advanced limestone and Dowa processes. Other raw material costs (adipic acid, formic acid, and aluminum sulfate) are minor.

The largest component of the first-year annual revenue requirements (as shown in Table S-4 for the base case application) for all processes is capital charges which are over 50% of the total. With the exception of raw material cost, all processes have the same order of major cost components: levelized capital charges, maintenance, plant and administrative overheads, and electricity, in order of decreasing importance. Raw material cost, however, is the second highest component of annual revenue requirements for the Davy S-H process, whereas, it is the fifth highest component for the advanced limestone and Dowa processes.

TABLE S-3. ANNUAL REVENUE REQUIREMENT SUMMARY

	Advanced limestone				Davy S-H				Dowa			
	First year		Levelized		First year		Levelized		First year		Levelized	
	M$	Mills/kWh	M$	Mills/kWh	M$	Mills/kWh	M$	Mills/kWh	M$	Mills/kWh	M$	Mills/kWh
New units[a]												
200 MW, 3.5% S	15.3	13.95	21.9	19.86	17.7	16.06	25.1	22.83	18.0	16.33	25.2	22.94
500 MW, 2.0% S	21.6	7.86	30.4	11.07	26.7	9.69	37.3	13.55	26.8	9.76	37.2	13.54
500 MW, 3.5% S	25.8	9.38	36.6	13.31	32.7	11.88	46.5	16.93	32.3	11.74	45.1	16.39
500 MW, 5.0% S	27.7	10.07	39.5	14.36	35.9	13.08	52.0	18.90	34.8	12.65	48.6	17.68
1,000 MW, 3.5% S	42.8	7.78	60.4	10.99	57.4	10.44	81.4	14.80	56.1	10.21	77.7	14.13
Existing units[b]												
200 MW, 3.5% S	17.1	15.53	22.0	20.04	21.9	19.87	27.8	25.28	22.7	20.66	28.7	26.14
500 MW, 3.5% S	32.5	11.83	41.5	15.10	40.2	14.63	51.2	18.62	40.9	14.86	51.4	18.71
1,000 MW, 3.5% S	56.9	10.34	72.1	13.12	71.6	13.02	90.6	16.48	71.9	13.08	90.0	16.36

a. New FGD facilities are constructed simultaneously with the power plant and have a 30-year remaining life.

b. Existing FGD facilities are retrofit installations on existing power plant facilities and have a 20-year remaining life.

TABLE S-4. MAJOR COST COMPONENTS FOR BASE CASE ANNUAL REVENUE REQUIREMENTS

	Major operating cost components (% of first-year annual revenue requirements)				
Process	1	2	3	4	5
Advanced limestone	Levelized capital charges 52.7	Maintenance 14.7	Plant and adm. overheads 12.0	Electricity 6.6	Raw materials 4.3
Davy S-H	Levelized capital charges 52.0	Raw materials 14.1	Maintenance 10.1	Plant and adm. overheads 8.3	Electricity 6.7
Dowa	Levelized capital charges 55.3	Maintenance 14.3	Plant and adm. overheads 10.9	Electricity 6.2	Raw materials 4.3

Capital investments play a major role in the ranking of the annual revenue requirements because levelized capital charges are a direct function of total capital investment and maintenance costs are a function of direct investment. In addition, overheads are a function of maintenance costs.

Case Variations--

As shown previously in Table S-3, the levelized annual revenue requirements for new units increase 70% to 85% as the power unit size increases from 200 to 500 MW and 175% to 225% as the power unit size increases from 200 to 1,000 MW. For existing units, the costs increase 80% to 90% for the 200- to 500-MW power unit size increase and 215% to 230% for the 200- to 1,000-MW range.

Although the Davy S-H process has the highest base case levelized annual revenue requirements, this relationship is not true of all power unit sizes. At the 200-MW size, the levelized annual revenue requirements for the Dowa process are slightly higher than those of the Davy S-H process. As the power unit size increases, the levelized annual revenue requirements for the Davy S-H process increase more rapidly than those of the Dowa process. This is due to the relationship between the variable raw material costs and the fixed-percentage capital-related costs (maintenance and levelized capital charges). Since the Dowa process has the highest total capital investment in every case, its maintenance cost and capital charges are also highest, but the Davy S-H process has the highest raw material cost for all cases. As the power unit size increases, the effect of raw material cost surpasses the effects of maintenance cost and capital charges.

As the coal sulfur increases from 2.0% to 3.5%, the levelized annual revenue requirements increase 20% for the advanced limestone and Dowa processes and 25% for the Davy S-H process. As the coal sulfur increases from 2.0% to 5.0%, the increases are 30% and 40%. The levelized annual revenue requirements of the Davy S-H process are equivalent to the Dowa process for the 2.0% sulfur coal case, but they are higher for the 3.5% and 5.0% sulfur coal cases. As in the power unit size variation, this trend is due to the relationship of raw material costs and capital-related costs.

The sensitivity of the Dowa process economics to changes in the absorber L/G ratio was evaluated since the base case costs were calculated using a L/G ratio of 45 gal/kaft^3, whereas recent test work has indicated that a L/G ratio of about 80 gal/kaft^3 is required for 90% SO_2 removal. The higher L/G ratio increased the capital investment 4%, primarily because of the larger pumps and piping required, and increased the levelized annual revenue requirements about 5% because of the higher electricity cost and the higher levelized capital charges. The levelized annual revenue requirements for the Dowa process became slightly higher than those of the Davy S-H process, but the cost differences between the two processes remain small.

CONCLUSIONS

The capital investment relationships among the processes remain the same for all power unit sizes and coal sulfur contents. Landfill capital investment is a minor element of capital investment and is similar for all of the processes.

The capital investment and annual revenue requirements of the advanced limestone process are substantially lower (about one-fourth and one-tenth, respectively) than those of the Davy S-H and Dowa processes. This is largely the result of a more extensive use of partial bypass for the advanced limestone process, which is possible because the use of adipic acid allows scrubbing at 95% removal compared with 90% for the other processes. The capital investments and annual revenue requirements of the Davy S-H and Dowa processes do not differ appreciably; the Davy S-H process is slightly lower in capital investment and slightly higher in annual revenue requirements. The cost relationships of the three processes hold true for different power unit sizes and coal sulfur contents, with the exception that the Davy S-H process annual revenue requirements increase slightly more rapidly with increasing power unit size and coal sulfur content.

The Davy S-H process is most sensitive to raw material cost changes because lime is used as the ultimate absorbent. The processes using limestone have absorbent costs three to four times lower. Other raw material costs (adipic acid, formic acid, and aluminum sulfate) are insignificant.

Approximately doubling the L/G ratio (from 45 to 80 gal/kaft^3) increases the capital investment and annual revenue requirements of the Dowa process 4% to 5%, which has no large effect on the cost relationships of the processes. The use of a chloride purge in the Dowa process, which is not used in the other processes, has no material effect on the costs.

Landfill is a minor element of the capital investments and annual revenue requirements of all the processes.

ECONOMIC EVALUATION OF ADVANCED LIMESTONE, DAVY S-H, AND DOWA GYPSUM-PRODUCING FGD PROCESSES

INTRODUCTION

Since 1967, the Tennessee Valley Authority (TVA) has analyzed the economics of flue gas desulfurization (FGD) processes and related electric utility fossil-fuel power plant technology. Many of these studies have been sponsored by the U.S. Environmental Protection Agency (EPA) as part of its FGD research and development program. The studies are based on current U.S. utility power plant conditions and are designed to allow equitable process comparisons. During this period, both the regulatory standards governing sulfur oxides (SO_x) emissions and the FGD technology to meet the regulations have evolved. Similarly, TVA's economic evaluation procedures have evolved and broadened to adapt to legislative, technical, and economic changes.

This study is the fifth in a series begun in 1977 for EPA to evaluate the economics of advanced FGD processes which have promise of commercial development. The first four studies (1-4) examined such FGD processes as limestone, lime, double alkali, citrate, magnesia, aqueous carbonate, and Wellman-Lord.

This study evaluates the economics of three gypsum-producing FGD processes. For many of the previous studies, the limestone wet-scrubbing process with pond disposal of the sulfite sludge has served as the standard against which other FGD processes were compared. However, gypsum-producing processes with landfill disposal are receiving much attention, principally because a landfill provides a more environmentally sound disposal method and requires significantly less land than a sulfite sludge pond.

The gypsum-producing FGD processes examined in this study are the advanced limestone process (limestone scrubbing with adipic acid addition and forced oxidation), the Davy Saarberg-Holter (S-H) process, and the Dowa (aluminum sulfate) process. The advanced limestone process differs from the previous standard limestone process in that adipic acid is used to buffer the pH of the scrubbing slurry, resulting in increased SO_2 absorption and limestone utilization. The Davy S-H process is a calcium-based process using a clear alkaline scrubbing solution buffered with formic acid and using lime as the primary raw material. The Dowa process is a double-alkali process which uses an aluminum sulfate solution to absorb SO_2 while using limestone to precipitate gypsum and regenerate the aluminum sulfate solution. All three processes include forced oxidation with sparged air to produce gypsum.

The capital investment and revenue requirements are estimated for a base case new, north-central, 500-MW power unit burning a 3.5% sulfur eastern bituminous coal. In addition, economics for case variations such as plant size, coal sulfur content, and existing plants are examined. The sensitivity of these processes to variations in costs of raw materials, electricity, maintenance, and labor are also examined. The FGD design is based on control of SO_x emission to meet the new source performance standards (NSPS) promulgated in 1979 (5).

BACKGROUND

Since the inception of limestone- and lime-based FGD processes, disposal of FGD byproduct sludge has been a major concern. The earliest FGD development work, however, focused on immediate concerns such as the development of effective process designs and operating conditions and on the control of problems such as plugging and scaling. As effective designs and operating conditions evolved, attention shifted to the development of processes which produce a more acceptable waste byproduct than scrubber sludge. Since the development of forced-oxidation techniques is a major factor leading to the development of each of the scrubbing processes evaluated in this report, factors which enhance forced-oxidation technology will be discussed, evolving into the overall development of the three processes included in this evaluation.

Forced oxidation in the context of FGD consists of treatment of the absorbent liquid with air to oxidize the sulfite formed by absorption of SO_2. In most cases, the sulfate thus formed is removed by precipitation as gypsum. In some cases, the objective of this additional process step is the production of gypsum that can be sold or given away for manufacturing purposes. In other cases, particularly in the United States, the objective is the production of a more tractable waste.

Essentially complete oxidation to sulfate is readily accomplished in simple equipment at prescribed conditions of pH and air-liquid contact. In most cases, the oxidation vessel is a single tank that operates at atmospheric pressure and without temperature control. Air is usually introduced through a simple manifold. Under favorable conditions, a liquid retention time of several minutes and air rates that provide 1 to 2 lb atoms of oxygen/lb mol of SO_2 absorbed are sufficient for essentially complete oxidation. A low pH is essential for practical oxidation rates, however. The oxidation rate decreases rapidly above a pH of 4 to 5 and becomes impractical at higher pHs.

Since low pHs are inimical to efficient SO_2 absorption, many ingenious schemes have been devised to reconcile the low pH necessary for oxidation with the higher pH for efficient SO_2 absorption. The diversity of absorbents and process designs used in forced-oxidation processes reflects this reconciliation.

PROCESS EVOLUTION

Intentional oxidation of sulfite in early FGD studies was considered as a means of improving SO_2 absorption efficiency or producing a salable sulfuric acid. The work of Johnstone (6), Copson and Payne (7), and Tarbutton and others (8) typically reports studies in which water containing a metal catalyst was used as the scrubbing medium and the oxygen was supplied by the gas containing the SO_2. Although effective in demonstrating the concept, these studies did not lead to effective FGD processes or production of an acid of salable concentration.

Forced oxidation using compressed air was considered for the lime slurry FGD process developed in England in the 1930s by J. Howden & Company and Imperial Chemical Industries, Ltd. (9). The objectives in this case were to produce a chemically stable waste and, possibly, a salable gypsum. Forced oxidation by air sparging with a manganese catalyst was subsequently used in a variation of the process (10).

Forced oxidation by air sparging was a common feature of FGD processes developed in the 1960s in Japan and the Federal Republic of Germany. Among these were the Dowa and Davy S-H processes evaluated in this study. The impetus to develop gypsum-producing processes in these countries has been ascribed (11) to concerns about disposal of sulfite sludges and, particularly in Japan, to a potential market for the gypsum (12). In some cases, these were double-alkali processes or based on the double-alkali principle to obtain both efficient SO_2 removal and efficient oxidation conditions.

Several full-sized limestone forced-oxidation FGD systems are in operation or under construction at U.S. utility power plants (13). Most of these processes are summarized by Knight and others (14). All are variations of the limestone slurry scrubbing process using different absorber designs and oxidation schemes. The first commercial systems were placed in operation in 1976 and 1977 at the Sherburne County Generating Plant of the Northern States Power Company (15). In most cases, the gypsum produced in these systems is discarded, although marketing is intended for the gypsum produced in at least two cases, the systems provided by Research-Cottrell, Inc. (R-C), now under construction at the Big Bend Station of Tampa Electric Company and at the Muscatine (Iowa) Power and Water power plant (16).

PROCESS DESIGN VARIATIONS

A low pH is of primary importance in attaining efficient oxidation under practicable FGD conditions of residence time and air rate. The increased oxidation efficiency at low pH is usually ascribed to the higher concentration of HSO_3^- at lower pHs, which increases the total concentration of oxidizable sulfite species in solution (17). Other effects of low pH, such as increased ion pair formation, have also been suggested (18). A pH of about 4.5 or less during oxidation is usual in most forced-oxidation processes (19), although in some cases, satisfactory oxidation has been attained at higher pHs (20).

Since the low pH desirable for oxidation is inimical to efficient SO_2 absorption, various techniques have been used to reconcile efficient oxidation with efficient SO_2 removal. For processes in which a low absorbent pH is inherent, the L/G ratio may be increased to compensate for the lower SO_2 absorption efficiency. An extreme example is the Chiyoda 101 process (21) which uses a dilute sulfuric acid solution with a pH of 1 as the absorbent and a L/G ratio of 200 to 300 gal/kft^3 (20).

A second alternative is to isolate the absorption and oxidation stages so that both can be operated at optimum pHs. In its simplest form, typified by several Japanese processes (22), the absorber liquid is maintained at a pH of about 6 and a bleedstream is oxidized in a separate system after being acidified with sulfuric acid to a pH of about 4. In these cases, the use of sulfuric acid is an additional expense and increases the quantity of product.

The desired pH control can also be attained by the use of two or more absorber stages with independent liquid systems and with series flow of absorbent from stage to stage countercurrent to the path of the flue gas. The first stage (relative to flue gas flow) can be operated at a low pH suitable for efficient oxidation but with limited SO_2 removal efficiency and the flue gas can be efficiently scrubbed in the subsequent stage or stages. The familiar venturi-absorber combination can be used in this manner, typified by the R-C absorber design (23) and the venturi-spray tower system used at the Shawnee test facility (24). The Pullman Kellogg horizontal absorber is also a multistage design (14).

Single-stage systems can also be made to operate at conditions that provide liquid alkalinities adequate for SO_2 removal at the absorber liquid inlet and a low pH in the effluent for efficient oxidation. Efficient absorbent utilization, so that the absorber effluent is depleted in alkaline species, is necessary. The resulting low pH allows oxidation in the effluent hold tank or an in-loop oxidation vessel before the liquid is treated with makeup absorbent to precipitate gypsum and replenish the alkalinity. A reactive absorbent; buffers to enhance alkalinity and control pH; and absorbers of special design, alone or in combination, are possible approaches. The Davy S-H process and the Kureha process (25) both use a reactive buffered absorbent liquid. In addition, the Davy S-H process uses a proprietary absorber. The Chiyoda 121 process is essentially an in-loop oxidation process using a proprietary absorber design with ordinary limestone slurry (26).

The use of limestone slurry as the scrubbing medium presents particular problems in forced-oxidation processes because of the slow dissolution rate of the limestone and the limited solubility of $CaCO_3$. The chemistry of limestone FGD has been widely discussed in FGD literature, some of which provide considerable insight into the chemistry of the process (27-30). Normally a relatively high stoichiometry is required for high SO_2 removal efficiency and the system operates at a high degree of disequilibrium in terms of the $CaCO_3/Ca^{2+}/CO_3^{2-}/CO_2$ system. The excess limestone remaining in the spent absorbent continues to dissociate, eventually producing a pH of about 6 (31) even though the pH leaving the absorber is low enough for efficient oxidation. Because of this, oxidation of a bleedstream is not practical in processes using limestone slurry without buffers. Also, efficient in-loop oxidation with single-stage systems may require reduced stoichiometries with a corresponding loss of SO_2 absorption efficiency (24). Many of the forced-oxidation processes described above were devised to circumvent this difficulty. Of the several general methods, the use of additives is particularly attractive. Additives used in small quantities are relatively inexpensive and require little additional equipment compared with the equipment and pumping requirements of two-stage processes, for example (4).

OXIDATION CHEMISTRY

In contrast to many facets of the chemistry of FGD processes, the oxidation of sulfite in aqueous solution has been the subject of investigations in several fields of science and technology for many years. These studies have produced a vast literature dealing with the many complexities of the oxidation reaction under the various physical and chemical conditions in which it is of academic or practical interest. This literature provides extensive general information on the chemical and physical processes involved in forced oxidation applied to FGD.

The reaction itself has attracted considerable attention in fundamental chemistry as a classic chain reaction. Although the basic mechanism which involves electron transfer among a number of intermediate species is well defined, some details of the basic mechanism remain unclear (32,33). The chain reaction is initiated by a number of catalysts and broken by various inhibitors. This, and the complexity of the kinetics, has led to a considerable variety in the methods and procedures by which the reaction is studied and applied, as well as in the results and the conclusions drawn from them. Mass transfer in gas-liquid systems and atmospheric processes are two particular areas of study which have produced an extensive and diverse literature on sulfite oxidation.

Catalysts and inhibitors play a central and somewhat complicated role in sulfite oxidation chemistry. The role of catalysts in initiating the oxidation reaction has been well known for many years (6). The effect of inhibitors has been similarly recognized (34). Practical applications of the reaction such as mass transfer determinations almost universally employ a catalyst. Clearly defined effects of both catalysts and inhibitors are complicated, however, by the involved kinetics of the reaction, which render the effects dependent on the characteristics of the system (35,36), and by the low concentrations at which both are effective. The effectiveness of some catalysts and inhibitors at very low concentrations also raises the question of contamination in even closely controlled experiments (37,38).

Metals of the first transition series, particularly manganese, iron, cobalt, and copper, are the most commonly used catalysts. Nonmetals such as NO_2 (39) and Cl_2 (40) are also effective. The role of copper is ambiguous because of its varying effects under different conditions (35). Both manganese (41) and iron (42) have been used in deliberately catalyzed FGD processes.

Certain materials, particularly organic compounds such as alcohols (34) and other hydroxyl-bearing compounds (43) are known to strongly inhibit the oxidation reaction. Copper was also found to be an inhibitor under some conditions (6). Hudson (44) investigated the effects of various inhibitors, including adipic acid, at typical FGD conditions and found the effects of catalysts and inhibitors at least qualitatively counteractive.

Mass transfer studies usually involve contact of oxygen or an oxygen-containing gas with a sulfite solution containing a catalyst. Linek and Vacek (45) have recently summarized many of these studies. Danckwerts (46) and de Wall and Okeson (37) also discuss the method. The procedure has practical advantages but suffers from the complexity of the kinetics and the susceptibility of the results to experimental conditions. Extensive studies of the kinetics have been made under both homogeneous and heterogeneous conditions, but models based on fundamental principles have not been developed (45). Much of the data provided by these studies were obtained under conditions of concentrations, temperatures, and pHs far removed from those found in FGD applications. They do, however, provide considerable insight into the mass transfer mechanisms and the complex role of catalysts. Hudson (44) and Hudson and others (47) have discussed many of the earlier studies from the aspect of FGD forced oxidation.

Studies related to the oxidation of sulfites under atmospheric conditions are also a rich source of information. Many of these have been discussed in recent publications (48,49,50). In contrast to mass transfer studies, these studies frequently involve complex reactant combinations in dilute solutions. In some aspects, this makes them more typical of FGD conditions.

ADDITIVES

Additives that enhance the SO_2 absorption efficiency may function either to increase the concentration of dissolved alkalinity or to buffer the liquid in a pH range more favorable to SO_2 absorption. Cations forming more soluble sulfite and sulfate salts than calcium are examples of the former. The presence of these cations create equilibrium relationships that increase the SO_3^{2-} concentration and decrease the Ca^{2+} concentration. Magnesium has been the most widely investigated and used additive of this type (27,51-54). It is particularly attractive because it forms $Mg^{2+}SO_3^{2-}$ ion pairs which are a major portion of the increased alkalinity in magnesium-enhanced systems (17). Since it functions primarily as an equilibria modifying additive, however, the effect is susceptible to other dissolved species having counter effects. Chlorides, for example, have a counter effect (30).

Buffering additives, in contrast, are essentially independent of the liquid composition. Buffers effective in increasing SO_2 absorption efficiency have pK_as between the pK_a of $SO_2 \cdot H_2O$ ($\approx$2) and the pK_a of $CO_2 \cdot H_2O$ ($\approx$7). Those most effective should have pK_as near the mean pH of the gas-liquid interface and the bulk liquid, which in limestone processes is usually 3.5 to 4.5 (55). By reducing the pH change as SO_2 is absorbed, the rate of SO_2 absorption is increased. The buffer may also serve to maintain a more effective oxidation pH. The primary effect of buffers on forced oxidation is the increased SO_2 absorption efficiency they provide. High SO_2 removal efficiencies can be obtained at lower slurry pHs. The increased absorbent utilizations and lower pHs both favor more efficient oxidation.

In addition to the pK_a requirement, a practical buffer should be a readily available inexpensive material that is reasonably stable and nonvolatile under FGD conditions. These requirements are met by a number of organic acids (56). Rochelle and King (17) discuss the chemistry and practicality of several organic acids, including adipic acid, as limestone FGD buffers.

Adipic acid was evaluated by EPA at the Industrial Environmental Research Laboratory/Research Triangle Park (IERL/RTP) pilot plant (57) and subsequently received extensive evaluation at the Shawnee test facility. Several absorber, hold tank, and forced-oxidation configurations were evaluated. The results of these and related studies have been extensively reported (58-60). In general, adipic acid at concentrations up to 3,000 ppm greatly improved SO_2 absorption efficiency and limestone utilization. Forced oxidation and high chloride concentrations had little or no effect on the effectiveness of the adipic acid. The lower pHs made possible by the use of adipic acid made in-loop oxidation practical at high SO_2 removal efficiencies and also made bleedstream oxidation possible. Extensive full-scale testing of adipic acid in forced-oxidation limestone FGD was begun in 1980 at the Southwest Power Plant of Springfield (Missouri) City Utilities (61) and the Rickenbacker Air National Guard Base near Columbus, Ohio (62) with similarly favorable results.

Adipic acid [H_2Ad, representing $HOOC(CH_2)_4COOH$] is regarded primarily as a buffer, although the adipate species also contributes to the alkalinity of the liquid (17).

$$Ad^{2-} + H^+ \rightleftarrows HAd^- \qquad (1)$$

$$HAd^- + H^+ \rightleftarrows H_2Ad \qquad (2)$$

With pK_1 and pK_2 values of about 4.9 and 4.0 in solutions containing chlorides (55), the adipic acid is primarily in the dissociated states in the regenerated absorbent liquid. As the pH decreases with SO_2 absorption, the adipate species recombine with the hydrogen ions. The reactions do not differ from those of the other proton-exchange reactions involved in SO_2

absorption, but they occur in the pH range of the mean of the interface and bulk liquid pH, thus serving to reduce the decrease in pH that results from SO_2 absorption.

Loss of adipic acid in the waste and through degradation in the absorber liquid system is a disadvantage that has received considerable attention. Adipic acid is lost in the liquid portion of the waste, adsorbed on the solids, and occluded in the solids (63). Forced oxidation reduces the quantity lost in the waste by all three processes (63) but may increase the quantity loss through degradation (64). Degradation losses are reduced by increasing the concentration (59). The environmental effects of adipic acid and possible degradation products in the waste have been investigated (60) and are not regarded as potentially harmful.

As an alternative, a byproduct of the adipic acid manufacturing process might also serve as a less expensive buffering agent. The byproduct, called dibasic acid, consists of a mixture of organic acids. Dibasic acid was successfully substituted for adipic acid at the Springfield (Winors) Water, Light, and Power Department's Southwest unit 1.

Formic acid, HCOOH, is also a common industrial commodity appreciably cheaper than adipic acid on a weight basis. Its effectiveness, in terms of SO_2 removal enhancement, is slightly less than half that of adipic acid on a molar concentration basis (55). Formic acid was evaluated as an additive to enhance SO_2 absorption by Hatfield and Potts (56). It was as effective as acetic, benzoic, and other acids tested. With a pK_a of about 3.5, formic acid buffers in the desired pH range to enhance SO_2 absorption, particularly at high chloride concentrations (55). A major objection often cited (17,56) against formic acid and other relatively volatile buffers is the potential loss through volatilization. Formic acid has a boiling point of about 213°F, appreciably lower than other organic acids useful as buffering additives.

WASTE DISPOSAL

Normally the waste slurry from forced-oxidation processes contains in excess of 95% gypsum and can be dewatered to 85% solids without difficulty. The dewatering characteristics have been extensively discussed (14,24). Dewatered gypsum is a loose solid that is regarded as physically suitable for disposal without further physical or chemical treatment. One disposal method is the stacking method commonly used in the phosphate industry (65). Stacking consists of pumping the slurry directly onto a diked disposal site, using the gypsum to increase dike height as the stack grows. The liquid flows through the stack or to drains within it and is collected for reuse. In addition to the substantial body of information on this process in the phosphate industry, the method was evaluated with FGD gypsum at the Scholz Power Plant of the Gulf Power Company in a demonstration program sponsored by the Electric Power Research Institute (EPRI) (66).

TYPING GUIDE SHEET
EPA-287 (Cin.)
(4-76)

Gypsum disposal has been evaluated in EPA-sponsored tests at the Shawnee test facility (67). Systematic evaluations of gypsum disposal by conventional landfill methods, such as those of stabilized high-sulfite wastes (68), have not been reported. However, gypsum disposal is likely to present many of the same constraints for seepage and runoff control as stabilized sulfite waste disposal. Except that the sulfite content is low or negligible, the same components are present, including minor and trace elements of environmental concern (69). Although some gypsum tends to undergo in situ cementation (70), exposed gypsum stacks may form extensive crevices and cavities due to settlement and seepage (70) and suffer surface erosion (71). Conversely, from a structural standpoint gypsum has been generously described as ". . . the most forgiving of all waste products . . ." (65). As such, landfill methods of seepage control and surface maintenance suitable for stabilized sulfite wastes should be applicable to untreated gypsum wastes.

ADIPIC ACID-ENHANCED FORCED-OXIDATION LIMESTONE PROCESS

The limestone process used in this study incorporates a spray tower absorber with in-loop forced oxidation in the absorber effluent tank. The design is based on data from tests at the Shawnee test facility (24,58,59). The single-stage, in-loop forced-oxidation configuration is used instead of other configurations tested because it is the simplest design and because it represents an apparent industry trend toward single-stage spray tower absorbers. The in-loop forced oxidation is made practical by the use of adipic acid.

Incorporation of the adipic acid additive and forced oxidation into the limestone process does not alter the basic chemistry of the process. It does, however, alter the importance in SO_2 absorption of various elements of the chemistry. The basic processes remain the absorption, hydration, and hydrolysis of SO_2 and the reactions of these products with various basic species formed by the addition of $CaCO_3$. Some oxidation of the sulfite occurs in the absorber by absorption of O_2 from the flue gas and the rest is oxidized by air sparged into the absorber effluent tank before the liquid is recirculated to the absorber.

The absorption of SO_2 is represented by the gas-liquid equilibrium of SO_2 and the aqueous species collectively identified as sulfurous acid.

$$SO_2(g) \rightleftarrows SO_2 \cdot H_2O \qquad (3)$$

$$SO_2 \cdot H_2O \rightleftarrows HSO_3^- + H^+ \qquad (4)$$

$$HSO_3^- \rightleftarrows SO_3^{2-} + H^+ \qquad (5)$$

The corresponding carbonate equilibria, from the standpoint of limestone dissolution, are

$$CaCO_3(s) \rightleftarrows Ca^{2+} + CO_3^{2-} \tag{6}$$

$$CO_3^{2-} + H_2O \rightleftarrows HCO_3^- + OH^- \tag{7}$$

$$HCO_3^- + H_2O \rightleftarrows H_2CO_3 + OH^- \tag{8}$$

$$H_2CO_3 \rightleftarrows CO_2(g) + H_2O \tag{9}$$

In the pH range typical of limestone FGD systems, the important aqueous species are $SO_2 \cdot H_2O$, HSO_3^-, SO_3^{2-}, HCO_3^-, and H_2CO_3. In addition, several complexes and ion pairs such as $CaHCO_3^+$ and $Ca^{2+}SO_3^{2-}$ are present in significant quantities (28). Magnesium, whether present as an impurity or as an additive, also forms similar species (17). SO_3^{2-} may be provided primarily by the dissolution of $CaSO_3 \cdot 1/2H_2O$. The species basic to $SO_2 \cdot H_2O$ is called the total dissolved alkalinity. Typical reactions are:

$$SO_2 \cdot H_2O + SO_3^{2-} \rightleftarrows 2HSO_3^- \tag{10}$$

$$SO_2 \cdot H_2O + Ca^{2+}SO_3^{2-} \rightleftarrows Ca^{2+} + 2HSO_3^- \tag{11}$$

$$SO_2 \cdot H_2O + HCO_3^- \rightleftarrows HSO_3^- + H_2CO_3 \tag{12}$$

$$H_2CO_3 \rightleftarrows CO_2(g) + H_2O \tag{13}$$

Adipic acid, although effective primarily as a buffer (equations 1 and 2), also contributes to the dissolved alkalinity.

$$SO_2 \cdot H_2O + Ad^{2-} \rightleftarrows HSO_3^- + HAd^- \tag{14}$$

$$SO_2 \cdot H_2O + HAd^- \rightleftarrows HSO_3^- + H_2Ad \tag{15}$$

The chloride ion also plays a role in the process chemistry because of its relative abundance in most coal-fired FGD applications. In terms of the equilibrium paradigm, Cl^-, which forms a highly soluble calcium salt, increases the Ca^{2+} ion concentration, which in turn decreases the concentration of such species as SO_3^{2-} and CO_3^{2-}.

$$CaCl_2 \rightarrow Ca^{2+} + 2Cl^- \quad (16)$$

$$CaSO_3 \cdot 1/2H_2O \rightleftarrows Ca^{2+} + SO_3^{2-} + 1/2H_2O \quad (17)$$

$$CaCO_3 \rightleftarrows Ca^{2+} + CO_3^{2-} \quad (18)$$

As the Ca^{2+} concentration increases with increasing Cl^- concentration, the anions in equations 17 and 18 must decrease to satisfy equilibrium relationships.

Dissolution of $CaCO_3$ and $CaSO_3 \cdot 1/2H_2O$ in the absorber is important sources of dissolved alkalinity. Either or both may be important depending on the conditions. According to studies at the Shawnee test facility (67) at high limestone stoichiometries (over about 1.3 mols $CaCO_3$/mol SO_2 absorbed), dissolved alkalinity is mostly a result of limestone dissolution which occurs primarily in the absorber, while at low stoichiometries (1.1 or less mols $CaCO_3$/mol SO_2 absorbed) $CaSO_3 \cdot 1/2H_2O$ dissolution also contributes to dissolved alkalinity.

The increase in Ca^{2+} concentration depends on the relative proportions of $CaCO_3$ and $CaSO_3 \cdot 1/2H_2O$ dissolved. Twice as much alkalinity, and calcium, is produced by dissolution of $CaCO_3$ as by an equal molar quantity of $CaSO_3 \cdot 1/2H_2O$.

$$2SO_2 \cdot H_2O + CaCO_3 \rightleftarrows Ca^{2+} + 2HSO_3^- + CO_2(g) + H_2O \quad (19)$$

$$SO_2 \cdot H_2O + CaSO_3 \cdot 1/2H_2O \rightleftarrows Ca^{2+} + 2HSO_3^- + 1/2H_2O \quad (20)$$

For a given quantity of SO_2, dissolution of $CaSO_3 \cdot 1/2H_2O$ increases the Ca^{2+} concentration twice as rapidly as $CaCO_3$. Processes depending on $CaSO_3 \cdot 1/2H_2O$ dissolution are thus more likely to exceed critical saturation levels of $CaSO_3 \cdot 1/2H_2O$ and $CaSO_4 \cdot 2H_2O$ in the absorber, with resulting plugging and scaling.

Sulfite oxidation can be represented by the overall expressions

$$2HSO_3^- + O_2 \rightarrow 2SO_4^{2-} + 2H^+ \quad (21)$$

$$2HSO_3^- + 2HCO_3^- + O_2 \rightarrow 2SO_4^{2-} + CO_2(g) + H_2O \quad (22)$$

Equation 22 is consistent with the observed increase in pH during forced oxidation in limestone FGD systems. Borgwardt (72) suggests that stripping of CO_2 from the liquid by air sparging increases $CaCO_3$ dissolution, following the relationship of equations 6 through 9. CO_2 stripping has also been proposed (67) for the unimpaired SO_2 absorption efficiency of forced-oxidation limestone processes. The more efficient dissolution of $CaCO_3$ as shown in equation 17 compensates for the loss of alkalinity derived from $CaSO_3 \cdot 1/2H_2O$ dissolution as shown in equation 16. Removal of sulfite species has also been shown in modeling experiments (55) to increase SO_2 absorption through the relationships shown in equations 3 through 5.

Forced oxidation also reduces the level of $CaSO_4$ supersaturation in the absorbent liquid. In EPA evaluations (24,72), saturation levels of about 100% were observed, considerably below the supersaturation levels in similar evaluations without forced oxidation. The lower saturations are attributed to the abundant seed crystals present in the absorbent liquid (24).

DAVY S-H PROCESS

The Davy S-H process is a lime-based, formic-acid-buffered process using in-loop forced oxidation. The process was developed by the West German company Saarberg-Holter Umwelttecknic GmbH, a joint venture by Saarbergwerke AG and Holter GmbH (73,74). Saarbergwerke AG is a large energy company and hard coal producer while Holter GmbH is an engineering company with experience in pollution control. Davy Corporation Limited acquired the North American license for the process in 1978 and markets it under the name Davy S-H (73).

A few S-H FGD systems are currently being operated or installed in Germany (74,75). FGD studies were begun by Saarbergwerke in 1968. In 1974, a 12-MW equivalent (30,000 m^3/h) S-H pilot plant was placed in operation on an incinerator at Solingen. In the same year, a prototype 40-MW system was started up at Weiher II, a coal-fired utility boiler at Saarbrucken. In 1976, a pilot system was operated on gas from a sintering plant. In 1979, a 13,000-m^3/h system was placed in operation on Claus units at a Veba Chemie AG plant at Gelsenkirchen and a 175-MW system was placed in operation on Weiher III. A second 175-MW system is planned for the same power unit. Additional systems are under construction at the oil-fired 300-MW Bewag power plant in West Berlin and a coal-fired 230-MW power plant in Voelklingen. In both cases, all the flue gas will be treated. The gypsum from the Bewag system will be marketed.

The process uses no purge streams to remove dissolved solids except those removed in the liquid remaining in the filtered gypsum. Accumulation of dissolved chlorides has been attributed to important chemical effects (76), including increasing the concentration of calcium in solution. Sulfate and sulfite concentrations are correspondingly reduced. The absorbent liquid can be increased to a high pH, up to pH 11 if necessary (77), without precipitation of calcium compounds. The highly basic liquid is an efficient absorbent for the overall reaction of dissolved SO_2 with the hydroxide ion.

$$SO_2 \cdot H_2O + OH^- \rightleftarrows HSO_3^- + H_2O \qquad (23)$$

In the absence of buffering pairs such as CO_3^{2-}/HCO_3^-, the pH of the liquid rapidly decreases until the buffering action of the formic acid becomes effective in the 5.0 to 4.5 pH range.

$$CHCOO^- + H_3O^+ \rightleftarrows HCOOH + H_2O \qquad (24)$$

The pH further decreases to about 4, which is suitable for oxidation. The rapid pH decrease in the absorber also ensures that the SO_2 absorbed is primarily in the form HSO_3^- so that the concentration of SO_3^{2-} does not exceed $CaSO_3 \cdot 1/2H_2O$ solubility limits and create plugging problems. Since the absorbent liquid is not recirculated directly to the absorber, SO_4^{2-} is not precipitated in the absorber.

Mechanically, the Davy S-H process is characterized by specially designed components and specific equipment arrangements, aspects of which are patented (77). The absorber (called a Rotopart gas scrubber) consists of a series of vertical tubular ducts about 3 feet in diameter (77) and 20 feet long (76) mounted on a larger horizontal vessel (the Rotopart itself) that serves as an entrainment separator. Flue gas flows downward through the ducts. The absorbent liquid is sprayed cocurrently into the flue gas as it enters the ducts. The ducts contain internal "shedding rings" to return the wall film to the gas and increase turbulence. The shedding rings are essentially multiple venturis (77). The flue gas and entrained liquid flows from the ducts into the Rotopart where the entrained liquid is separated by deflection plates and a curved-path stripper jacket that acts as a centrifugal separator. The Rotopart absorber operates at a moderate flue gas velocity and pressure drop and a low L/G ratio (76). The unit is modular in that the capacity is determined by the number of vertical ducts rather than the diameter of the ducts. According to published information (77), the Weiher III unit was designed for a total 10-in. H_2O pressure drop (8-in. H_2O in the absorber ducts) at 12 ft/sec, and a L/G ratio of about 20 gal/kaft^3. L/G ratios of 10 gal/kaft^3 have also been cited (76).

The absorbent liquid flows by gravity into unagitated oxidation vessels with a height to diameter ratio greater than 1 and a retention time of about 30 minutes (77). Air is sparged into the liquid through pipes extending downward from the tops of vessels at stoichiometries up to 5 lb atoms O/lb mol SO_2 absorbed (78). The air serves in lieu of an agitator to mix the liquid and keep the gypsum formed in suspension. Oxidation tanks in series can be used to reduce air requirements (77).

The oxidized slurry flows by gravity from the oxidation tanks through a "mixing trough." Lime slurry and makeup formic acid are added to this trough. The lime slurry addition rate is based on pH control of the recovered liquid (77). The stoichiometric ratio is about 1.01 to 1.02 mols CaO/mol (SO_2 + 2HCl) absorbed (77). Published descriptions (75,79) cite the possible need of HCl addition if sufficient chloride is not provided by the flue gas.

The slurry flows from the mixing channel into a thickener consisting of two concentric sections. The thickener bottoms are dewatered in a conventional rotary vacuum filter and the filtrate is returned to the thickener. The thickener overflow is pumped to the Rotopart absorber. This is the only major liquid process flow in the system that is pumped.

DOWA PROCESS

The Dowa process is unique among operating wet FGD processes in using an aluminum-based absorbent. It is a double-alkali process incorporating a solution of basic aluminum sulfate as the absorbent, forced oxidation by air sparging in the absorber liquid loop, and precipitation of gypsum in the bleedstream by addition of pulverized limestone. The process was developed by the Dowa Mining Company of Japan in the 1960s where it was placed in commercial operation on a kiln in 1972 (80). Subsequently, by 1979, seven other installations were placed in operation (81). The process is vended in the United States by the Air Correction Division of Universal Oil Products, Inc. (UOP), as U.S. licensee. A 10-MW scale evaluation of the process was conducted in 1980 at the Shawnee test facility (82) in a project jointly funded by TVA, EPRI, and UOP (83). Another test at the same facility is planned in 1982. The attractiveness of the process for U.S. applications lies primarily in the use of limestone rather than lime as the precipitating agent.

A basic aluminum sulfate process was developed by Imperial Chemical Industries (ICI) in England in the 1930s. It was a regeneration process designed to recover SO_2 from smelter gases (84). The basic aluminum sulfate solution was prepared by treating a 90- to 100-g/L aluminum sulfate solution with limestone and filtering out the gypsum formed. The absorbed SO_2 was evolved from the scrubbing liquid by heating. A pilot plant was operated in England and a commercial unit was operated in Finland in the 1930s. A small plant was reported operating in England in the 1950s. In all cases, an SO_2-rich smelter gas was treated. Johnstone and others (85) regarded the process as favorable in terms of the heat required to evolve the SO_2 but considered basic aluminum sulfate solutions unsuitable for treating dilute gases ($<1\%$ SO_2) because of their low absorption capacity.

The eight Dowa processes operating in Japan in 1978 scrubbed about 1,400,000 ft^3/min of gas containing 600 to 7,500 ppm of SO_2 (81). One unit is on an oil-fired boiler; the others are on metal kilns and sulfuric acid plants. In most cases, the nonboiler units treat gas somewhat cooler than utility boiler flue gas. Mobile-bed or packed absorbers, gas velocities of about 5 ft/sec, and L/G ratios of 20 to 75 gal/kaft^3 have been reported (80,86) for these units. The boiler installation treats flue gas at 338°F containing 1,500 ppm SO_2. A packed tower with a gas velocity of 4 ft/sec and a L/G ratio of 75 gal/kaft^3 is used (81). All of the Dowa installations in Japan attain over 90% SO_2 removal.

The Dowa process operates in a pH range of 3 to 4. This promotes oxidation and a high limestone utilization efficiency, although to the detriment of SO_2 removal efficiency, which is compensated for by a L/G ratio somewhat higher than double-alkali processes that operate at higher pHs (12).

The efficacy of aluminum in enhancing SO_2 absorption lies in the strong tendency of hydrated aluminum ions to hydrolyze. In its simplest form, the general reaction end members are

$$[Al(H_2O)_6]^{3+} + H_2O \rightleftarrows [Al(OH)(H_2O)_5]^{2+} + H_3O^+ \qquad (25)$$

$$[Al(OH)_2(H_2O)_4]^{+} + H_2O \rightleftarrows [Al(OH)_3(H_2O)_3] + H_3O^+ \qquad (26)$$

in which the hexaquo aluminum ion becomes successively hydrolyzed until equilibrium is reached. For the initial hydrolysis, the approximate pK_a is about 5 (87), showing that the hydrolysis is extensive and accounting for the acidic nature of the solution. The reactions shown are too simple in that the actual ionic species are a complex mixture of hexacoordinated mononuclear and polynuclear species such as $[Al_2(OH)_2]^{4+}$ and $[Al_{13}(OH)_{32}]^{7+}$, both of which have been identified (88). In the presence of sulfate ions, similar basic aluminum sulfate species form, their nature depending on the conditions of the solution. Several dozen of these have been identified (89).

A disadvantage of the use of aluminum is the extremely insoluble nature of the hydroxide, which has a Ksp of about 10^{-33}. To avoid precipitation of $Al(OH)_3$, a low pH and a low aluminum concentration must be maintained, both of which limit the SO_2 absorption efficiency of the solution. This is offset to some extent by the strong buffering nature of the various hydrated and hydrolyzed aluminum species, which occurs in the pH range of about 3.5 to 4.5.

The term basicity is used to define the degree to which the aluminum ion is hydrolyzed. A basicity of zero can be represented by the form $[Al(H_2O)_6]^{3+}$ and a basicity of 100% can be represented by the form $[Al(OH)_3(H_2O)_3]$. This range of basicities occurs within the acidic pH range. In the basic pH range, basicity increases above 100%, culminating in the destruction of the polynuclear hydrated complexes in which the canonical form $Al(OH)_3$ exists as a solid, and the resolution of the aluminum as an aluminate, $[Al(OH)_4(H_2O)_2]^-$. The basicities of interest in the Dowa process lie below 40% at aluminum concentrations of 15 to 20 g/L, conditions which avoid precipitation of aluminum from the solution (82).

Basicity is controlled by addition of a suitable proton acceptor, A.

$$[Al(H_2O)_6]^{3+} + A \rightleftarrows [Al(OH)(H_2O)_5]^{2+} + HA \qquad (27)$$

In the Dowa process, this is $CaCO_3$, which also serves to precipitate gypsum.

$$2[Al(H_2O)_6]^{3+} + CaCO_3 \rightleftharpoons Ca^{2+} + 2[Al(OH)(H_2O)_5]^{2+} + CO_2(g) + 2H_2O \qquad (28)$$

$$SO_4^{2-} + Ca^{2+} \rightarrow CaSO_4.2H_2O(s) \qquad (29)$$

The few published discussions of the Dowa process (80,82) follow the only primary publication (86) in using the forms $Al_2(SO_4)_3$, $Al_2(SO_4)_3.Al_2O_3$, and $Al(OH)_3$ to represent 0%, 50%, and 100% basicity, respectively. Thus the absorption stage is represented by

$$Al_2(SO_4)_3 \cdot Al_2O_3 + 3SO_2 \rightarrow Al_2(SO_4)_3 \cdot Al_2(SO_3)_3 \qquad (30)$$

which provides a measure of stoichiometric unity. The essential nature of the chemistry, however, lies in the nature of the dissolved aluminum and sulfite ions.

The overall chemistry of the Dowa process can be represented in simplified form as occurring in three stages.

1. Absorption of SO_2 into a solution of basic $Al_2(SO_4)_3$, in which the primary active species are the partially hydrolyzed and hydrated aluminum ions, which act as proton acceptors, and the dissolved SO_2, which acts as a proton donor.

$$SO_2(g) + H_2O(l) \rightleftharpoons SO_2 \cdot H_2O \qquad (31)$$

$$SO_2 \cdot H_2O + H_2O \rightleftharpoons HSO_3^- + H_3O^+ \qquad (32)$$

$$[Al(OH)(H_2O)_6]^{2+} + H_3O^+ \rightleftharpoons [Al(H_2O)_6]^{3+} + 2H_2O \qquad (33)$$

   Some carbonate species are also present because of the $CaCO_3$ added in the regeneration stage and CO_2 present in the flue gas. The extent to which these species contribute to SO_2 removal or to buffering is not quantified. The low concentration of sulfite species in the entering absorbent may also make a significant contribution to SO_2 removal.

2. Oxidation in an oxidation tank in the absorber liquid system using a conventional air sparging system. Essentially complete oxidation of the sulfite species occurs, enhanced by the low pH of the solution.

3. Regeneration of a bleedstream by addition of a controlled quantity of pulverized limestone. The aluminum ions are rehydrolyzed and the sulfate is precipitated as gypsum.

The quantity of limestone added must be carefully controlled to produce the desired basicity without precipitation of aluminum hydroxides and sulfates (82). Since the concentration of carbonate species in the absorber liquid probably plays only a secondary role in SO_2 removal, and since limestone dissolution is rapid at the pH of the system, limestone stoichiometry is essentially 1 mol of $CaCO_3$/mol of SO_2 absorbed.

The solids are removed by conventional dewatering techniques and the solution is returned to the absorber liquid system. Aluminum lost in the solids is replaced by addition of $Al_2(SO_4)_3$. An additional process step may be necessary to control chloride and (if gypsum for wallboard manufacture is desired) magnesium concentrations. In this case, a purge stream is treated with sufficient limestone to precipitate $Al(OH)_3$, which is recovered, and the solute, containing chlorides and magnesium, is discarded.

Commercial installations of the Dowa process in Japan differ in several design features (80,86). Either packed or mobile-bed absorbers are used along with design variations in the neutralizing and dewatering system. Both mobile-bed and packed absorber configurations were evaluated in tests at the Shawnee test facility (82). The process is more precisely characterized by the chemistry of the process than by a particular design philosophy.

Typically, the process consists of a single-pass absorber containing mobile or fixed packing. The absorbent liquid is recirculated from a tank beneath the absorber, through an oxidizer vessel, and back to the absorber. L/G ratios reported in commercial units range up to about 75 gal/kaft^3 at a gas velocity of about 4 ft/sec (90). An aluminum concentration of about 20 g/L is maintained by addition of Al_2SO_3 solution to the absorber liquid loop.

A bleedstream is withdrawn from the absorber liquid loop and passes through a redissolution tank (see below) to one or two neutralizing tanks to which pulverized limestone is added to precipitate gypsum and regenerate the absorbent. The slurry from the neutralization tank is thickened and dewatered in a conventional thickener-centrifuge or thickener-filter system and the recovered absorbent is returned to the absorber liquid loop.

To control the chloride concentration of the absorbent liquid (and the magnesium concentration if the gypsum is to be used for wallboard manufacture), a purge stream is withdrawn from the bleedstream and treated with excess pulverized limestone. A gel of mixed aluminum salts and gypsum is precipitated, leaving a magnesium and chloride solution which is disposed of. The gel is returned to the redissolution tank where the aluminum salts dissolve.

A distinctive feature of the Dowa process is the close control of the absorbent liquid alkalinity (86). The basicity of absorbent liquid rather than the pH is the basis of control. Because of the buffering action, the absorbent pH varies very little over a wide range of basicities. The basicity of the absorbent entering the absorber and the basicity of the liquid overflow from the neutralizing tank are continuously analyzed in an automatic analyzer. The basicity of the absorber liquid loop is controlled at about 10% by the rate at which the absorbent liquid is withdrawn and regenerated absorbent liquid added. The basicity of the neutralization tank overflow is controlled at about 30% by the limestone addition rate. Close control of both is necessary to obtain the maximum SO_2 absorption efficiency and to avoid precipitation of aluminum salts.

CENTER OF PAGE
TOP OF IMAGE AREA

DESIGN AND ECONOMIC PREMISES

The premises used in this study have been developed by TVA and EPA during similar economic evaluations made by TVA since 1967. The design premises are formulated to quantify SO_2 emissions from a representative electric utility coal-fired power plant boiler and to establish efficiencies, process flow rates, and other design and operating conditions to meet environmental regulations. The economic premises are designed to represent the many factors affecting costs. They prescribe procedures for determining capital investment and annual revenue requirements based on regulated utility economics. Results are reported in sufficient detail to permit comparisons with previous economic studies and to permit adjustments based on use of alternative premises. The costs are based on 1982 costs for capital investment and 1984 costs for annual revenue requirements.

The economic evaluations are based on a conceptual design developed from the design premises and engineering data such as flow diagrams, material balances, and equipment costs. A base case new 500-MW power unit burning a 3.5% sulfur, 16% ash bituminous coal (dry basis), and complying with the 1979 NSPS (5) is used as the primary basis of comparison. Case variations in coal sulfur content and power unit size and for retrofit applications illustrate their effects on the economics of the processes evaluated.

DESIGN PREMISES

Coal Premises

The coals used are three eastern bituminous coals containing 5.0%, 3.5%, and 2.0% sulfur, all on a dry basis. The assumed properties of these coals are based on composites of many samples representing major coal production areas (91,92). The eastern bituminous coals are assumed to have a heating value of 11,700 Btu/lb and an ash content of approximately 15.1% as fired. The composition for the coals is shown in Table 1.

Power Plant

The power plant site is assumed to be in the east-north-central region of the United States (Illinois, Indiana, Ohio, Michigan, and Wisconsin). The location represents an area in which coal-fired power plants burning coals of diverse type and source are situated (93,94). The design is based on standard design practices (95,96) and current trends in utility boiler construction (97,98). The base case power unit is a new, single 500-MW, balanced-draft, horizontally fired, dry-bottom boiler burning pulverized coal. Heat rates,

Preceding page blank

PAGE NUMBER
BOTTOM OF IMAGE AREA OUTSIDE DIMENSION FOR TABLES AND ILLUSTRATIONS
TYPING GUIDE SHEET

TABLE 1. COAL COMPOSITIONS

						Ultimate analysis				
Coal	Sulfur, %		Ash, %	Moisture, %	Heat content, Btu/lb	C, %	H, %	O, %	N, %	Cl, %
	Dry basis	As-fired basis								
Eastern bituminous	5.0	4.80	15.10	4.0	11,700	65.2	4.0	5.5	1.3	0.1
Eastern bituminous	3.5	3.36	15.14	4.0	11,700	66.7	3.8	5.6	1.3	0.1
Eastern bituminous	2.0	1.92	15.08	4.0	11,700	67.8	3.7	6.0	1.4	0.1
		Moisture-free basis								
Eastern bituminous	5.0	5.00	15.7			67.9	4.2	5.7	1.4	0.1
Eastern bituminous	3.5	3.50	15.7			69.5	4.0	5.8	1.4	0.1
Eastern bituminous	2.0	2.00	15.7			70.6	3.9	6.3	1.4	0.1

shown in Table 2, are based on power unit size. To provide for equitable comparisons, the power units are not derated for energy consumption by the FGD systems evaluated. Instead, the energy requirements are charged as independently purchased commodities. Cost estimates are based on a single power unit independent of other units at the site. Power unit size case variations consist of similar 200- and 1,000-MW units. The FGD systems being evaluated as new units are assumed to be installed during construction of the power plant and are assumed to have a 30-year life and to operate at full load for 5,500 hours per year. This operating schedule is equivalent to a total lifetime operation of 165,000 hours. Identical existing power units with 20 years of remaining life at 5,500 hours per year of full-load operation are also evaluated. This operating schedule is equivalent to a total remaining lifetime operation of 110,000 hours.

TABLE 2. POWER UNIT OPERATING TIME AND HEAT RATE

	New			Existing		
Power unit size, MW:	200	500	1,000	200	500	1,000
Remaining life, years	30	30	30	20	20	20
Full load, hr/yr	5,500	5,500	5,500	5,500	5,500	5,500
Heat rate, Btu/kWh	9,700	9,500	9,200	9,900	9,700	9,500

Flue Gas Compositions

Flue gas compositions are based on combustion of pulverized coal, assuming a total air rate equivalent to 139% of the stoichiometric requirement (defined as air for combustion of carbon, hydrogen, and sulfur). This includes 20% excess air to the boiler and 19% additional air leakage to the flue gas in the air heater. It is assumed that 80% of the ash present in each coal is emitted as fly ash and 92% of the sulfur is emitted as SO_x. The remaining sulfur is removed in the bottom ash and fly ash. No loss of sulfur in the pulverizers is assumed. Three percent of the sulfur emitted as SO_x is SO_3 and the remainder is SO_2. Flue gas compositions and flow rates based on combustion of each of the coals assumed for the study are shown in Tables 3 through 5.

Environmental Regulations

Emissions from new coal-fired utility plants are regulated by the NSPS, which are issued under authority of Section 111 of the Clean Air Act as amended in 1970 and 1977. This section requires EPA to set Federal emission limitations that reflect the degree of control that can be achieved by application of the best system of continuous emission reduction. On

TABLE 3. FLUE GAS COMPOSITION

FOR 5.0% SULFUR EASTERN BITUMINOUS COAL

(Gas from air heater to electrostatic precipitator)

Component	Volume, %		Lb-mol/hr	Lb/hr
N_2	75.13		136,900	3,834,000
O_2	5.53		10,080	322,600
CO_2	12.10		22,040	970,000
SO_2	0.30	(2,976 ppm)	542	34,750
SO_3	0.01	(93 ppm)	17	1,343
NO	0.03	(324 ppm)	59	1,766
NO_2	0.00	(16 ppm)	3	142
HCl	0.01	(66 ppm)	12	418
H_2O	6.89		12,550	226,000
	100.00		182,200	5,391,000
Fly ash				50,290
Total				5,441,000

Sft3/min (60°F) = 1,152,000
Aft3/min (300°F) = 1,684,000

Fly ash loading

	Gr/sft3
Wet	5.09
Dry	5.47

Sulfuric acid dewpoint temperature: 316°F

TABLE 4. FLUE GAS COMPOSITION

FOR 3.5% SULFUR EASTERN BITUMINOUS COAL

(Gas from air heater to electrostatic precipitator)

Component	Volume, %	Lb-mol/hr	Lb/hr
N_2	75.22	137,500	3,852,000
O_2	5.54	10,130	324,200
CO_2	12.33	22,550	992,300
SO_2	0.21 (2,079 ppm)	380	24,320
SO_3	0.01 (66 ppm)	12	940
NO	0.03 (324 ppm)	59	1,766
NO_2	0.00 (16 ppm)	3	142
HCl	0.01 (66 ppm)	12	418
H_2O	6.65	12,160	219,100
	100.00	182,800	5,415,000
Fly ash			50,050
Total			5,465,000

Sft3/min (60°F) = 1,156,000
Aft3/min (300°F) = 1,690,000

Fly ash loading

	Gr/sft3
Wet	5.05
Dry	5.41

Sulfuric acid dewpoint temperature: 308°F

TABLE 5. FLUE GAS COMPOSITION

FOR 2.0% SULFUR EASTERN BITUMINOUS COAL

(Gas from air heater to electrostatic precipitator)

Component	Volume, %		Lb-mol/hr	Lb/hr
N_2	75.24		137,800	3,860,000
O_2	5.54		10,100	324,800
CO_2	12.52		22,920	1,009,000
SO_2	0.12	(1,185 ppm)	217	13,900
SO_3	0.00	(38 ppm)	7	537
NO	0.03	(324 ppm)	59	1,766
NO_2	0.00	(16 ppm)	3	142
HCl	0.01	(66 ppm)	12	418
H_2O	6.54		11,970	215,600
	100.00		183,100	5,426,000
Fly ash				49,480
Total				5,475,000

Sft3/min (60°F) = 1,158,000
Aft3/min (300°F) = 1,692,000

Fly ash loading

	Gr/sft3
Wet	4.98
Dry	5.33

Sulfuric acid dewpoint temperature: 297°F

December 23, 1971, EPA issued NSPS to limit emissions of SO_2, NO_x, and particulate matter from utility power plants (99). In 1979, EPA revised the NSPS (5), which are shown in Table 6. The controlled outlet SO_2 emission and SO_2 removal efficiencies for premise coals are shown in Figure 1 and tabulated in Table 7. Existing units are not restricted by the revised NSPS but are regulated by individual state implementation plans (SIP) promulgated in accordance with the Clean Air Act. Due to the variability and site-specific nature of SIP and for comparison purposes, the SIP which governs existing units in this report are assumed to be equal to the revised NSPS emission standard of 0.6 lb SO_2/MBtu heat input.

TABLE 6. 1979 REVISED NSPS EMISSION STANDARDS

SO_2[a]

70% SO_2 removal (minimum) to a maximum SO_2 emission of 0.6 lb SO_2/MBtu
0.6 lb SO_2/MBtu maximum emission up to 90% SO_2 removal
90% SO_2 removal (minimum) to a maximum SO_2 emission of 1.2 lb SO_2/MBtu
1.2 lb SO_2/MBtu maximum emission

Particulate[a]

0.03 lb/MBtu

Source: (5)
a. All quantities are based on heat input.

TABLE 7. PREMISE COAL EMISSION STANDARDS

Coal	Equivalent SO_2 content of coal, lb SO_2/MBtu	Overall equivalent SO_2 removal efficiency, %	Equivalent SO_2 removal required in FGD system, %[a]	Controlled outlet emission, lb SO_2/MBtu
Eastern bit., 5.0% S	8.21	90.0	89.1	0.82
Eastern bit., 3.5% S	5.74	89.6	88.7	0.60
Eastern bit., 2.0% S	3.28	81.7	80.1	0.60

a. Based on FGD system as the only SO_2 control device and the previously defined sulfur retention in the ash.

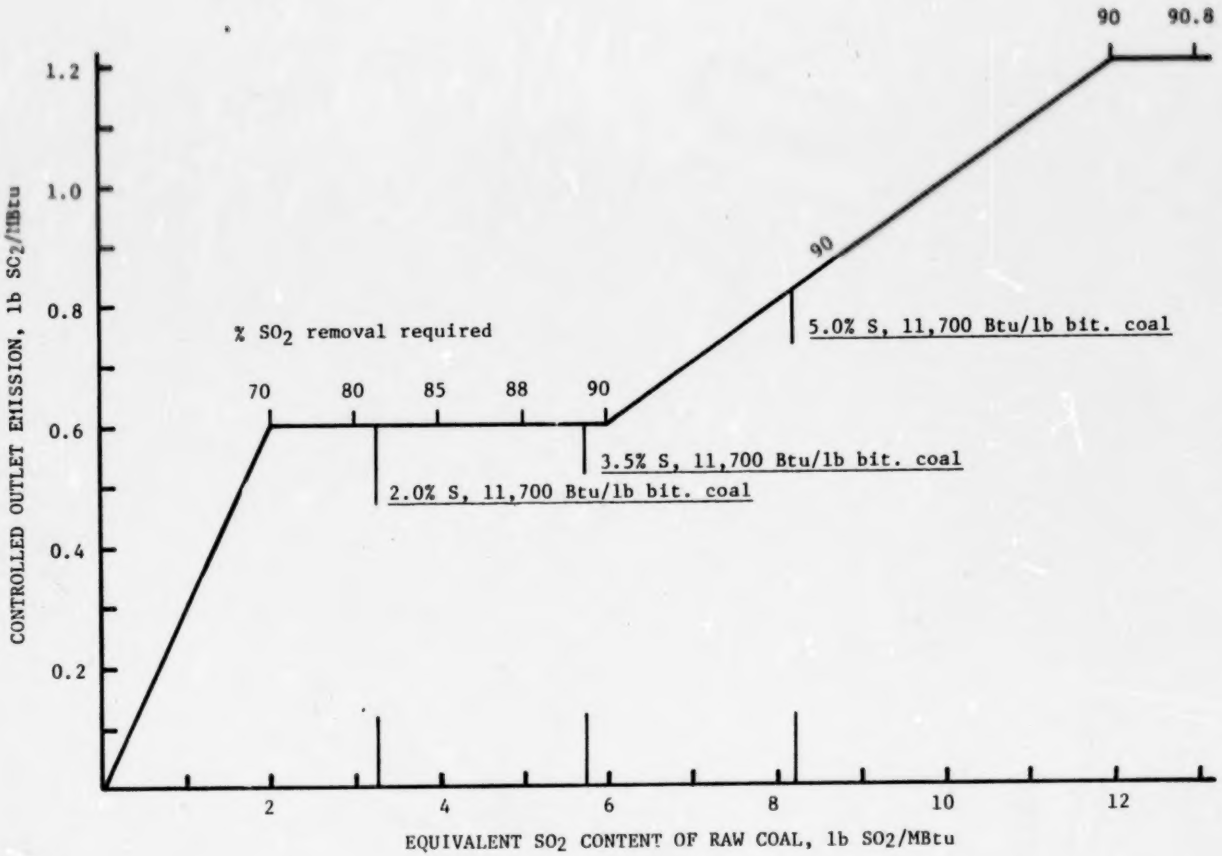


Figure 1. Controlled SO_2 emission requirements for 1979 NSPS. Premise coals, shown underlined, are based on premise boiler conditions.

Particulate Matter

Cold-side (post-air-heater) ESPs sized to meet the 0.03 lb/MBtu heat input emission standard are assumed for particulate matter control. The costs for ash collection and disposal are not included in the economic evaluations.

Flue Gas Desulfurization

The conceptual design of the FGD system meets applicable emission standards and reflects a practical operating approach. FGD systems are close coupled to the power unit by a plenum into which the power plant induced-draft (ID) fans discharge. The plenum allows the FGD systems to be designed for a different number of trains than the number of power plant ducts (to account for limitations in the available size of individual absorber units), and it facilitates the use of spare trains. To minimize flow control problems which can result from this design, separate fans are provided in each train. Conventional power plant ID fans operating balanced draft in respect to the boiler are assumed upstream from the plenum to overcome the pressure drop of the boiler. Since they are required even if FGD units are not installed, the installation and operating costs are not included in the costs of the FGD system. Separate ID fans are provided in each FGD train downstream from the reheater to overcome the pressure drop created as a result of installing FGD facilities.

The FGD costs include FGD-related ductwork and associated equipment between the power plant ID fans and the stack plenum. All ductwork between the power plant ID fan and the stack plenum is charged to the flue gas treatment system. This is done on the assumption that without the flue gas treatment system, the boiler ID fans would discharge directly into the stack plenum. Absorber trains are sized for a maximum of 125 MW of flue gas up to a maximum flue gas rate of 513,000 sft^3/min (60°F). Thus, the 500-MW base case requires four operating trains and the 200- and 1,000-MW case variations require two (100 MW) and eight (125 MW) operating trains, respectively. It is assumed that the annual availability of an absorber train is 85% and that no scrubbing time is lost during startup. Spare absorber trains are provided as described below.

Emergency Bypass--

Because the 1979 NSPS allow emergency bypass around the FGD system under some conditions if spare scrubbing capacity is provided, redundancy in the form of spare trains and provision for bypass of 50% of the gas that would normally be scrubbed are included in all FGD economic evaluations. The 1,000-MW case variation with eight operating trains is provided with two spare trains. Smaller boilers are provided with one spare train. An emergency bypass of 50% of the scrubbed gas is assumed to be an economic balance between the higher cost of providing additional bypass and the small likelihood of multiple absorber train failures making higher bypass rates necessary. The bypass is installed as two identical ducts from each end of the inlet plenum to the stack plenum downstream from the absorber trains.

Partial Scrubbing--

Partial scrubbing is used for all of the advanced limestone process cases. The use of adipic acid results in 95% SO_2 removal efficiency in the limestone absorber and it is more economical to partially scrub at 95% SO_2 removal efficiency and bypass the remaining flue gas to minimize reheat requirements. For the Davy S-H and Dowa processes, partial scrubbing at a 90% SO_2 removal efficiency and combining it with the remaining flue gas is more economical only for the 2.0% sulfur coal. In these partial scrubbing cases, the bypassed gas duct requirements and the emergency bypass capability are combined in the same duct. The ducts are sized to handle both the flue gas normally bypassed and the emergency bypass of 50% of the gas normally scrubbed. The amount bypassed varies from 6% to 16%, depending on the process and the sulfur in coal as shown below.

Coal	Flue gas bypass, % Advanced limestone	Davy S-H	Dowa
Eastern bit., 5.0% S	6.2	-	-
Eastern bit., 3.5% S	6.7	-	-
Eastern bit., 2.0% S	15.7	11.0	11.0

Ductwork--

Square ductwork with 2-inch insulation is used for the inlet plenum and absorber trains. To prevent ash settling, a gas velocity of 50 ft/sec is used for the inlet plenum, all ductwork, and the emergency bypass. A gas velocity of 25 ft/sec is used for the reheater section. Duct material is 3/16-inch Cor-Ten steel when the gas temperature is higher than 150°F and 3/16-inch stainless steel when the gas temperature is lower than 150°F.

Removal Efficiencies--

It is assumed that 50% of the SO_3, 95% of the HCl, none of the NO_x, and 50% of the remaining fly ash in the flue gas are removed in the FGD system.

Spare Equipment--

Equipment is spared in accordance with general field practice. This includes a spare train of crushing and grinding equipment, slakers, waste filters, pumps, and the spare scrubbing train or trains.

Mist Eliminator--

The mist eliminator is a three-pass, open-vane chevron type. The mist eliminator reduces entrained moisture to a maximum level of 0.1% (by weight) of the flue gas. This maximum level is assumed for calculation of the amount of stack gas reheat required.

Stack Gas Reheat--

Indirect steam reheat to 175°F is provided to evaporate entrained water droplets not removed by the mist eliminator and to increase plume buoyancy.

Approximately one-half of the reheater tubes are made of Inconel 625 and approximately one-half of Cor-Ten. Inconel 625 is highly resistant to corrosion and is used for the first bank of tubes, which increases the flue gas temperature to 150°F. The succeeding banks of Cor-Ten tubes raise the temperature of the gas at the exit to 175°F. For cases with partial bypass, the scrubbed gas is heated to a temperature less than 175°F because of the heat obtained from the bypassed gas. In these cases, the total amount of reheater surface decreases; however, the percentage of Inconel 625 tubes may increase to as much as 100% (for reheat to 150°F or less).

Raw Materials and Byproducts--

Raw materials and byproduct storage capacity is 30 days. Standard raw material characteristics are shown in Table 8.

Solids Disposal

The solids generated by the processes evaluated are disposed of by landfill at a site one mile from the FGD facilities. Sufficient land is provided for disposal for the projected life of the FGD facility. Disposal costs for fly ash is not included. The disposal site is assumed to be an area of low relief with sufficient soil for landfill requirements.

Landfill--

Square area-type landfills with a 20-foot-high perimeter and a 6-degree cap are used, as shown in Figure 2. After topsoil removal, the landfill area is lined with 12 inches of clay (assumed available onsite). A French drain system (perforated pipe surrounded by gravel) in 24 inches of bottom ash is placed on top of the clay. The bottom ash layer and French drain system allow the seepage to be collected separately from runoff and treated for pH adjustment.

Land requirements include the landfill, catchment basin, equipment storage area, topsoil storage area, and a 50-foot perimeter of undisturbed land. Costs for access roads; a 6-foot security fence around the landfill area and the catchment basin; security lighting; and topsoil stripping, replacement, and revegetation are included. One upstream and three downstream groundwater monitoring wells are also included.

ECONOMIC PREMISES

Schedule and Cost Factors

A 3-year construction period, from early 1981 to late 1983, is used. Mid-1982 costs are used for capital investment. Mid-1984 costs are used for annual revenue requirements. These costs represent the midpoint of construction expenditures in 1982 and the midpoint of the first year of operation in 1984. Costs are projected from Chemical Engineering cost indexes (100), as shown in Table 9. Frequently used costs are shown in Table 10.

TABLE 8. RAW MATERIAL CHARACTERISTICS

	Size as received	Ground size	Analysis[a]	Bulk density, lb/ft3
Limestone	0 x 1-1/2 inch	90% to pass 325 mesh	95% $CaCO_3$ 0.15% MgO 4.85% inerts 5 lb H_2O/100 lb dry limestone	95
	Fineness of grind index factor = 5.7 Hardness of work index factor = 10			
Lime (pebble)	3/4 x 1-1/4 inch	-	95% CaO 0.15% MgO 4.85% inerts 5 lb H_2O/100 lb dry lime	55
Adipic acid	Crystalline powder	-	99.8% $(CH_2)_4(COOH)_2$ 0.2% inerts	49
Formic acid	90% HCOOH	-	90% HCOOH 10% H_2O	-
Ferric chloride	30% aqueous solution	-	30% $FeCl_3$ 70% H_2O	-
Anti-foulant	Aqueous solution	-	Nalco 7319	-
Aluminum sulfate	45% to pass 35 mesh, 8% to pass 200 mesh	-	83% $Al_2SO_4 \cdot 18H_2O$ 17% Al_2O_3	60

a. Limestone and lime analysis on a dry basis. H_2O is based on pounds of dry limestone or lime.

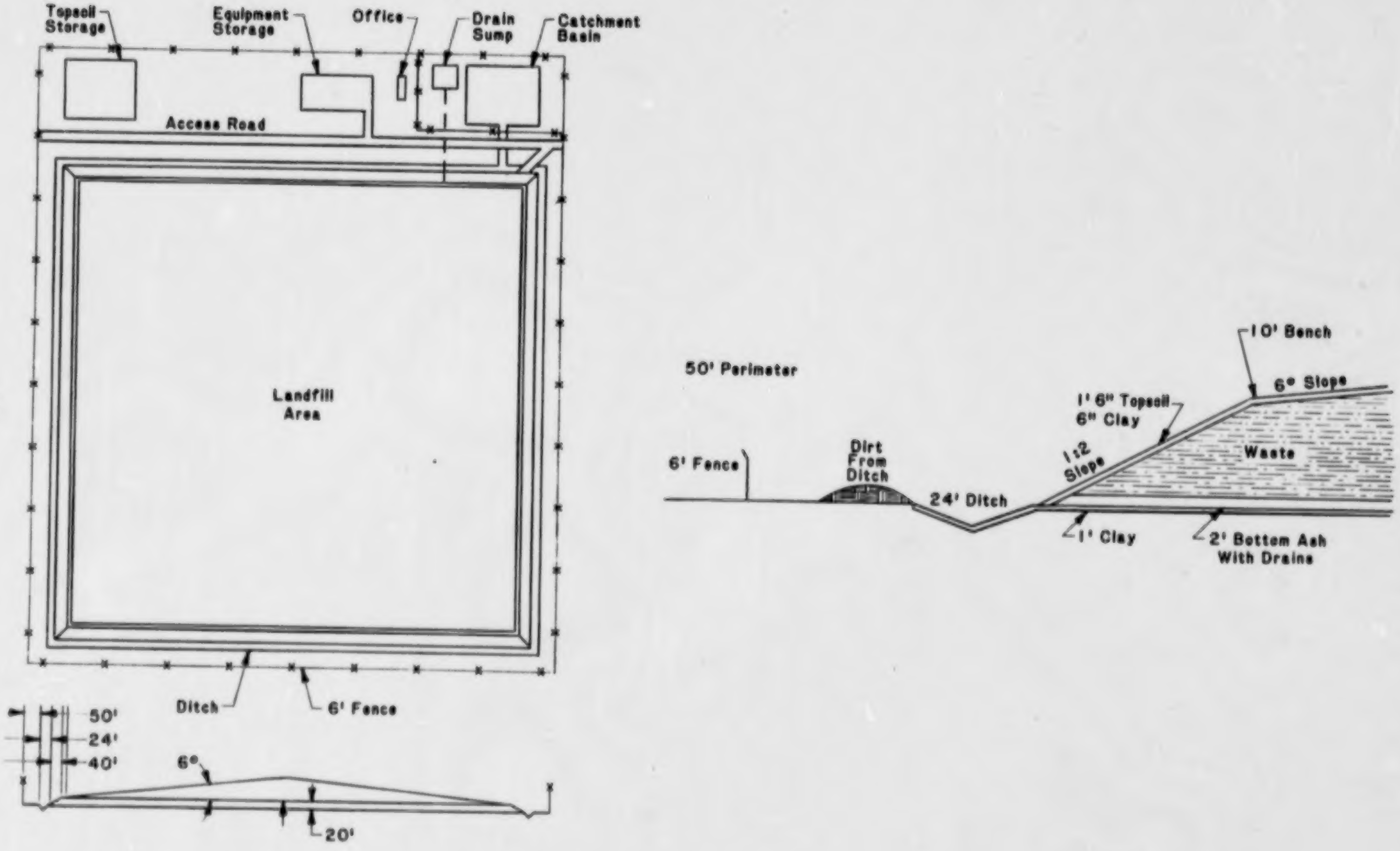

Topsoil Storage
Equipment Storage
Office
Drain Sump
Catchment Basin
Access Road
Landfill Area
Ditch
6' Fence
50'
24'
40'
6°
20'
50' Perimeter
6' Fence
Dirt From Ditch
24' Ditch
1:2 Slope
1' 6" Topsoil
6" Clay
10' Bench
6° Slope
Waste
1' Clay
2' Bottom Ash With Drains

Figure 2. Landfill plan and construction details.

TABLE 10. COST FACTORS

Project Timing	
Construction	
Start	January 1981
End	December 1983
Midpoint	Mid-1982
First-year operation	Mid-1984
1984 Utility Costs	
Electricity	$0.037/kWh
Steam	$2.50/klb, $3.30/MBtu
Diesel fuel[a]	$1.60/gal
Filtered river water	$0.14/kgal (up to 0.6 Ggal)
	$0.12/kgal (0.6-2 Ggal)
	$0.10/kgal (2-5 Ggal)
	$0.08/kgal (over 5 Ggal)
1984 Labor Costs	
Operating labor - FGD	$15.00/man-hr
- Waste disposal	$21.00/man-hr
- Analyses	$21.00/man-hr
1984 Raw Material Costs	
Limestone	$8.50/ton (95% $CaCO_3$, dry basis)
Lime	$75.00/ton (pebble 95% CaO, dry basis)
Adipic acid	$1,500/ton
Formic acid	$760/ton
Ferric chloride	$234/ton (anhydrous)
Antifoulant	$2,840/ton
Aluminum sulfate	$350/ton

These cost factors are based on a rural east-north-central plant location.

---

[a] Cost is based on wholesale price of barge-load quantities. Road taxes are not included.

TABLE 9. COST INDEXES AND PROJECTIONS

Year:	1978	1979	1980	1981[a]	1982[a]	1983[a]	1984[a]
Plant	218.8	238.7	261.1	277.1	297.9	320.2	342.6
Material[b]	240.6	264.4	292.6	311.2	336.1	363.0	388.4
Labor[c]	185.9	194.9	204.3	227.3	245.5	265.2	283.7

a. TVA projections (these projections continue to be used to maintain consistency with previous and ongoing studies).
b. Same as "equipment, machinery, supports" Chemical Engineering index.
c. Same as "construction labor" Chemical Engineering index.

Capital Cost Estimates

The capital investment estimates are divided into three major sections: direct investment, indirect investment, and other capital investment. The sum of the direct and indirect investments is the fixed investment. Costs included in each of these sections are discussed below.

Direct Investment--

Direct investment consists of the installed costs of all process equipment, including provisions for services, utilities, and miscellaneous; and waste disposal investment. Installation costs are estimated by major processing area and include charges for all piping, foundations, excavations, structural steel, electrical equipment, instruments, ductwork, paint, buildings, taxes, freight, and a premium for 7% overtime construction labor.

Service facilities such as maintenance shops, stores, communications, security, offices, and road and railroad facilities are estimated or allocated on the basis of process requirements. Included in the utilities investment are necessary electrical substations, conduit, steam, process water, fire and service water, instrument air, chilled water, inert gas, and compressed air distribution facilities. Services, utilities, and miscellaneous are estimated as 6% of the total process capital.

All equipment and direct construction costs associated with waste disposal are included in landfill costs. All mobile equipment involved in loading and transporting the waste from the in-process storage area, as well as working the landfill, is included in landfill equipment cost. The landfill construction cost is listed separately from the landfill equipment. The sum of total process capital; services, utilities, and miscellaneous; and the landfill costs is the total direct investment.

Indirect Investment--

Indirect capital costs cover fees for engineering design and supervision, architect and engineering contractor, construction expense, contractor fees, and contingency. Listed in Tables 11 and 12 are the percentages used to calculate these investment costs.

TABLE 11. INDIRECT CAPITAL COST FACTORS

	Process[a]			Landfill[b]		
	1,000 MW	500 MW	200 MW	1,000 MW	500 MW	200 MW
Engineering design and supervision	6	7	8	2	2	2
Architectural and engineering contractor	1	2	3	1	1	1
Construction expense	14	16	18	7	8	9
Contractor fees	4	5	6	4	5	6
Total	25	30	35	14	16	18

a. Percentage of process direct investment, excluding landfill.
b. Percentage of landfill equipment and construction.

TABLE 12. CONTINGENCY AND ALLOWANCE FOR STARTUP AND MODIFICATIONS COST FACTORS

	Contingency, % of direct and indirect investment		Allowance for startup and modifications, % of fixed investment	
Process	Process[a]	Landfill[b]	Process[a]	Landfill[b]
Advanced limestone	10	20	8	0
Davy S-H	20	20	10	0
Dowa	20	20	10	0

a. Percentage of only process costs.
b. Percentage of only landfill costs.

Other Capital Investment--

The allowance for startup and modifications is applied as a percentage of the total fixed investment. Since the startup and modification costs for the landfill area are assumed to be negligible, this allowance is calculated using only the process fixed investment. The percentages used to calculate these costs are listed in Table 12.

The cost of borrowed funds (interest) during construction is 15.6% of the fixed investment (both process and waste disposal). This factor is based on an assumed 3-year construction schedule and is calculated using a 10% weighted cost of capital, with 25% of the construction expenditures in the first year, 50% in the second year, and 25% in the third year of the project construction schedule. Expenditures in a given year are assumed uniform over that year. Startup costs are assumed to occur late enough in the project schedule that there are no charges for the use of money to pay startup costs. No royalty charges are included for the advanced limestone and Dowa processes. Royalty charges for the Davy S-H process were obtained from the process vendor.

All land associated with the process and waste disposal area is charged to the process. The cost of land is $5,000 per acre.

Working capital is the amount of money invested in raw materials, supplies, finished products, accounts receivable, and money on deposit for payment of operating expenses. Working capital is defined as the equivalent cost of one month's raw material cost, 1.5 months' conversion cost, 1.5 months' plant and administrative overhead costs (all of the above are shown in the annual revenue requirements sheet), and 3% of the total direct capital investment. One month is defined as 1/12 of annual costs.

Retrofit Factor--

For existing plant cases, a retrofit factor of 1.3 is assigned to cover the additional investment required for fitting the equipment into available space. Each of the area investments (e.g., materials handling, etc.) is multiplied by the retrofit factor.

Annual Revenue Requirements

Annual revenue requirements account for recovery of various direct and indirect operating and maintenance costs and capital charges. Annual revenue requirements normally vary from year to year as operating and maintenance costs change and capital charges decline. Thus, no single year is necessarily representative of the lifetime costs, and single-year undistorted comparisons cannot be made among processes with different ratios of operating costs to capital charges. In addition, it is necessary to take into account the effect of time on the value of money (i.e., for inflation, the future earning power of money spent, and other factors).

Frequently these factors are accounted for by levelizing (101). Levelization converts all the varying annual revenue requirements to a constant annual value, such that the sum of the present worths of the levelized annual revenue requirements equals the sum of the present worths of the actual variable annual revenue requirements. The levelized value is calculated by multiplying the revenue requirements for each year by the appropriate present worth factor and summing the present worth values. Then the single present worth value is converted to equal annual values by multiplying the result by the capital recovery factor.

In these premises, the operating and maintenance costs are levelized by multiplying the first-year operating and maintenance cost by a levelizing factor. The levelized capital charges are determined by levelizing the percentage of capital investment applied yearly as capital charges. The levelizing factor includes a discount factor reflecting the time-value of money and an inflation factor reflecting the effects of inflation during the operating life of the system. The discount rate used is 10% and the annual inflation rate used is 6%. The levelizing factor also depends on the remaining life of the system.

Operating and Maintenance Costs--

Operating and maintenance costs consist of direct costs for raw materials and conversion and indirect costs for overheads. Conversion costs consist of labor and supervision, utilities, maintenance, and analysis costs.

Raw materials include consumables required for their chemical or physical properties, other than fuel for the production of heat. In this evaluation, the raw materials consist of limestone, lime, adipic acid, formic acid, ferric chloride, an antifoulant, and aluminum sulfate. Raw material costs are determined from vendor quotations or published sources. All costs are delivered costs.

Operating labor and supervision consists of all labor requirements for operation of the FGD-related equipment. The allocation of operating labor and supervision man-hours depends on the process complexity, number of process areas, labor intensity of the process, and operating experience.

Services such as steam, electricity, process water, and fuel are charged under the utilities heading. Costs for steam and electricity are based on the assumption that the required energy is purchased from another source. This simplifying assumption eliminates the need to derate the utility plant. Process water requirements are defined as any water used by the process being evaluated and are determined from the material balance. Steam requirements are for stack gas reheat. Electrical power requirements are determined from the installed horsepower of operating electrical equipment (excluding the horsepower of spared equipment). Each motor in operation is assumed to be operating at rated capacity, although this results in higher power consumptions than would actually occur. Electrical requirements are obtained from the equipment list where the motor horsepower is identified, plus an additional amount for functions such as lighting.

Process maintenance costs are calculated as a percentage of direct process investment which varies with the process complexity, process equipment, materials handled, process areas, and power unit size. For the advanced limestone and Dowa processes, the percentage is 9% for the 200-MW cases, 8% for the 500-MW cases, and 7% for the 1,000-MW cases. For the Davy S-H process, the percentages are 7%, 6%, and 5%, respectively. Landfill maintenance costs are calculated as 3% of the total landfill disposal investment.

Analysis costs are based on process complexity and are listed as a single entry.

Plant and administrative overheads include plant services, general engineering (excluding maintenance), interplant communications and transportation, and the expenses connected with management activities. Plant and administrative overheads for the FGD process are calculated as 60% of the conversion costs less utilities.

Capital Charges--

Capital charges are those costs incurred by construction of the facility that must be recovered during its life. They consist of returns on equity and debt (discount rate), depreciation, income taxes, and other costs such as insurance and local taxes. In keeping with common practice for investor-owned utilities, the weighted cost of capital is used as the discount rate (102). Depreciation is stated as a sinking fund factor to simplify calculations. An allowance for interim replacement is included to compensate for possible early retirement of the facility. Credits are also included for tax preference allowances. The capital charges are shown below.

	% of total capital investment	
	Remaining life, years	
	30 (new plant)	20 (existing plant)
Weighted cost of capital	10.00	10.00
Depreciation (sinking fund factor)	0.61	1.75
Annual interim replacement	0.56	0.67
Levelized accelerated tax depreciation	-1.36	-1.43
Levelized investment tax credit	-1.93	-2.14
Levelized income tax	4.31	4.08
Insurance and property taxes	2.50	2.50
Levelized annual capital charge	14.7	15.4

The capital charges are applied as a percentage of the total capital investment, including land and working capital. Although land and most of working capital cannot be depreciated and are not subject to investment tax credit, their inclusion has an insignificant effect on capital charges.

The capital structure is assumed to be 35% common stock, 15% preferred stock, and 50% long-term debt. The cost of capital is assumed to be 11.4% for common stock, 10.0% for preferred stock, and 9.0% for long-term debt. The weighted cost of capital is 10.0%. The discount rate is equal to the weighted cost of capital. Other economic factors used in financial calculations are a 10% investment tax credit rate, 50% State plus Federal income taxes, 2.5% property tax and insurance, and an annual inflation rate of 6%. Salvage value is assumed to be less than 10% and equal to removal cost.

The sinking fund factor method of depreciation is used since it is equivalent to straight-line depreciation levelized for the economic life of the facility using the weighted cost of capital. The use of the sinking fund factor does not suggest that regulated utilities continue to use sinking fund depreciation. An annual interim replacement (retirement dispersion) allowance is also included as an adjustment to the depreciation account to ensure that the initial investment will be recovered within the actual rather than the forecasted life of the facility. Tax preference allowances are incentives designed to encourage investment as a stimulus to the overall economy. The basic accounting method used is the flow-through method which passes the tax advantage to revenue requirements as soon as they occur.

Levelized Operating and Maintenance Costs--

Assuming a constant inflation rate, the levelized operating and maintenance costs are determined by multiplying the first-year operating and maintenance costs by an appropriate levelizing factor, L_f. The levelizing factor is calculated as follows:

$$L_f = CRF_e\ (K + K^2 + K^3 + --- + K^N)$$

$$= CRF_e\ [K(1 - K^N)]/(1 - K)$$

where: CRF_e = capital recovery factor for the economic life
$K = (1 + i)/(1 + r)$; present worth of an inflationary value
i = inflation rate
r = discount rate
N = book life in years

An inflation rate of 6% ($i = 0.06$) and a discount rate of 10% ($d = 0.10$) are used for all units. The first-year operating and maintenance costs are multiplied by the levelizing factor to obtain the levelized operating and maintenance costs.

ACCURACY OF ESTIMATES

The accuracy of the preliminary-grade estimates used in this evaluation is -15%, +30%. It represents the potential variation in costs for an actual installation in comparison to the estimated costs, expressed as a percent of the estimated cost. The accuracy assigned to a cost estimate is empirical and not related to variabilities in a statistical sense. Rather, it depends on both the quantity and the quality of the technical data available. Accuracy ranges also reflect the numerous uncertainties surrounding estimates made using simplifying assumptions. However, when comparing the preliminary-grade costs for processes evaluated using the same methodology, many of the same simplifying assumptions are made for each of the processes. Therefore, the comparability is greater than the overall accuracy of the estimates. When directly comparing estimates of the same grade, the uncertainty ranges associated with the compared costs are estimated to be less than ±10%.

SYSTEMS ESTIMATED

The base case conceptual designs are prepared from process descriptions, material balances, flow and control diagrams, design drawings, and equipment requirements. These are developed from vendor information, industry experience, and the premises described in the previous section. Each process is divided into an area-by-area format to facilitate the comparison of equitable processing steps. This format is used in the following system descriptions, equipment lists, and capital investment tables.

In all cases, upstream removal of fly ash to meet the 1979 NSPS of 0.03 lb/MBtu heat input is assumed. Fly ash control costs are not included in the FGD costs. FGD costs begin with a plenum downstream of the power plant ID fans that distributes the flue gas to the individual absorber trains. The absorber trains discharge the flue gas into the stack plenum which is not included in the FGD costs.

ADVANCED LIMESTONE PROCESS

The advanced limestone process included in this study incorporates adipic acid addition to enhance SO_2 absorption and forced oxidation to produce gypsum. SO_2 is absorbed into a limestone slurry in a countercurrent spray tower and the absorber effluent is oxidized to gypsum in an in-loop tank by sparging it with air. Gypsum crystals are concentrated in a thickener and further dewatered by filters. The gypsum cake is disposed of in a landfill. The advanced limestone process flow diagram is shown in Figure 3 and the base case material balance is shown in Table 13. A plan and elevation drawing and a plot plan are shown in Figures 4 and 5. The equipment list, itemized by area, and the purchase cost of the equipment are shown in Table 14. The costs do not include installation.

Major Processing Areas

The advanced limestone process is divided into seven processing areas: materials handling, feed preparation, gas handling, SO_2 absorption, stack gas reheat, oxidation, and solids separation and disposal.

Materials Handling--

Limestone, crushed and screened to 0 x 1-1/2 inches at the quarry and containing 95% $CaCO_3$, is delivered by truck or rail and unloaded by conveyor to a 30-day stockpile. Facilities also include conveyors and elevators to transport limestone approximately 150 feet from the stockpile to the limestone preparation area. Adipic acid is delivered by trucks, which are unloaded pneumatically, and is stored in a 30-day storage silo.

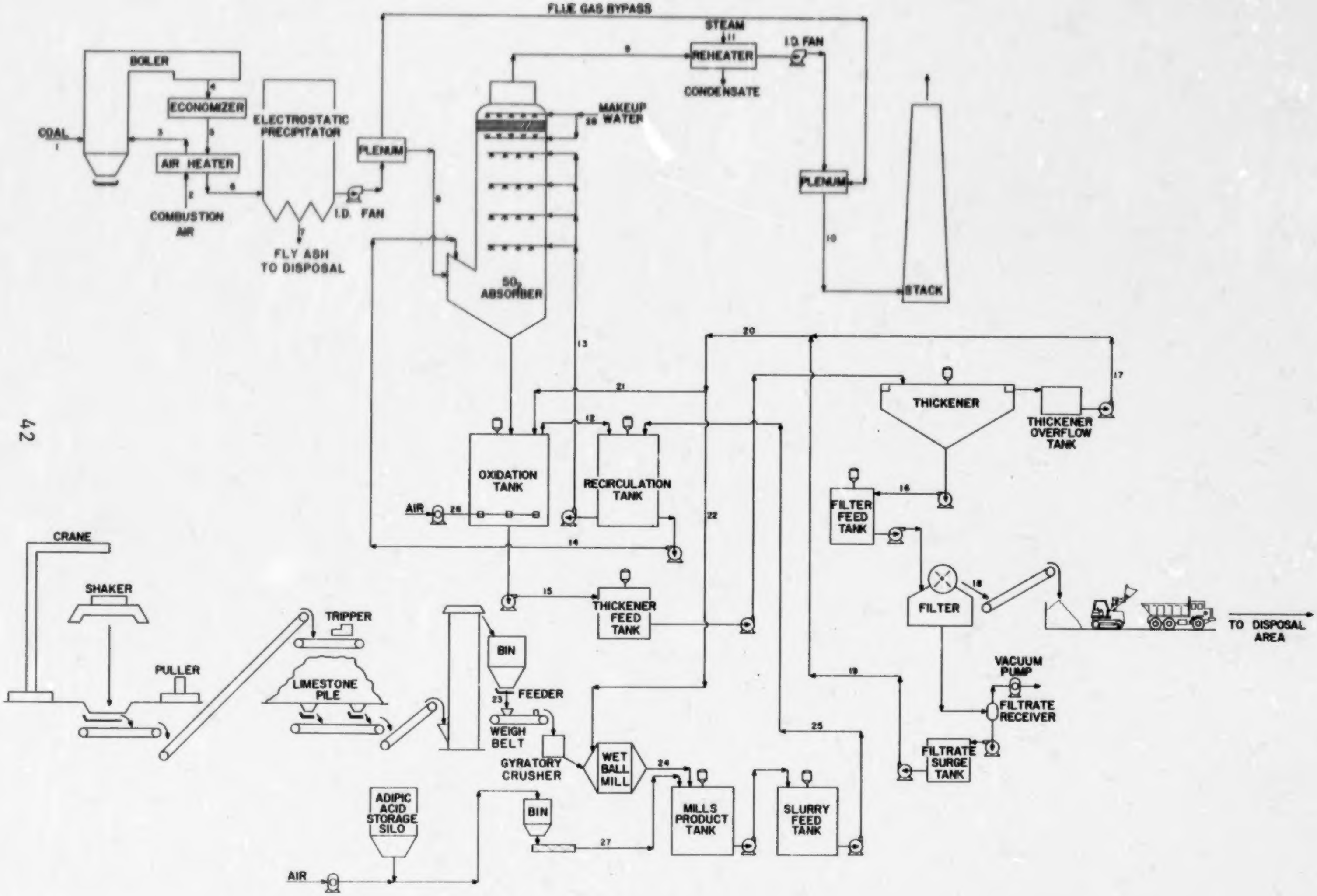

FLUE GAS BYPASS
BOILER
COAL
1
ECONOMIZER
AIR HEATER
COMBUSTION AIR
2
3
4
5
6
ELECTROSTATIC PRECIPITATOR
FLY ASH TO DISPOSAL
7
I.D. FAN
PLENUM
8
9
SO2 ABSORBER
MAKEUP WATER
28
STEAM
11
REHEATER
CONDENSATE
I.D. FAN
PLENUM
10
STACK
13
20
21
12
OXIDATION TANK
RECIRCULATION TANK
AIR
26
14
22
THICKENER
THICKENER OVERFLOW TANK
17
16
FILTER FEED TANK
FILTER
18
TO DISPOSAL AREA
VACUUM PUMP
FILTRATE RECEIVER
FILTRATE SURGE TANK
19
CRANE
SHAKER
PULLER
TRIPPER
LIMESTONE PILE
BIN
FEEDER
23
WEIGH BELT
GYRATORY CRUSHER
15
THICKENER FEED TANK
WET BALL MILL
24
25
MILLS PRODUCT TANK
SLURRY FEED TANK
ADIPIC ACID STORAGE SILO
BIN
27
AIR

Figure 3. Advanced limestone process flow diagram.

TABLE 13. ADVANCED LIMESTONE PROCESS

MATERIAL BALANCE

Stream No.		1	2	3	4	5
Description		Coal to boiler	Total air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,072,000	4,378,000	4,772,000	4,772,000
2						
3	Flow rate, sft^3/min(60°F)		1,120,000	967,300	1,003,000	1,003,000
4	Temperature, °F		80	495		705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

Stream No.		6	7	8	9	10
Description		Gas to electrostatic precipitator	Fly ash to disposal	Gas to presaturator	Gas to reheater	Gas to stack
1	Total stream, lb/hr	5,465,000	50,050	5,051,000	5,260,000	5,623,000
2						
3	Flow rate, sft^3/min(60°F)	1,156,000		1,077,000	1,152,000	1,231,000
4	Temperature, °F	300		300	124	175
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

Stream No.		11	12	13	14	15
Description		Steam to reheater	Oxidation tank overflow	Recycle slurry to absorber	Recycle slurry to presaturator	Absorber slurry to thickener
1	Total stream, lb/hr	77,300	57,054,000	54,400,000	2,720,000	794,900
2						
3	Flow rate, sft^3/min(60°F)					
4	Temperature, °F	470				
5	Pressure, psig	500				
6	Flow rate, gpm		108,700	103,500	5,175	1,512
7	Specific gravity		1.05	1.05	1.05	1.05
8	pH					
9	Undissolved solids, %		8	8	8	8
10						

Stream No.		16	17	18	19	20
Description		Thickener bottoms to filter	Thickener overflow return	Gypsum cake to disposal	Filtrate return	Clear liquor return
1	Total stream, lb/hr	159,000	612,900	74,810	84,160	697,100
2						
3	Flow rate, sft^3/min(60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm	240	1,226		168	1,395
7	Specific gravity	1.3	1.0	2.0	1.0	1.0
8	pH					
9	Undissolved solids, %	40		85		
10						

(Continued)

TABLE 13. (Continued)

	Stream No.	21	22	23	24	25
	Description	Clear liquor to oxidation tank	Clear liquor to wet ball mill	Limestone to weigh feeder	Slurry to mills product tank	Limestone slurry to recirculation tank
1	Total stream, lb/hr	672,100	24,940	41,110	66,050	66,100
2						
3	Flow rate, sft^3/min(60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm	1,345	50		83	83
7	Specific gravity	1.0	1.0		1.6	1.6
8	pH					
9	Undissolved solids, %				60	60
10						

	Stream No.	26	27	28		
	Description	Air to oxidizer	Adipic acid to mills product tank	Makeup water to absorber		
1	Total stream, lb/hr	59,950	51	264,900		
2						
3	Flow rate, sft^3/min(60°F)	13,060				
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm			530		
7	Specific gravity			1.0		
8	pH					
9	Undissolved solids, %					
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

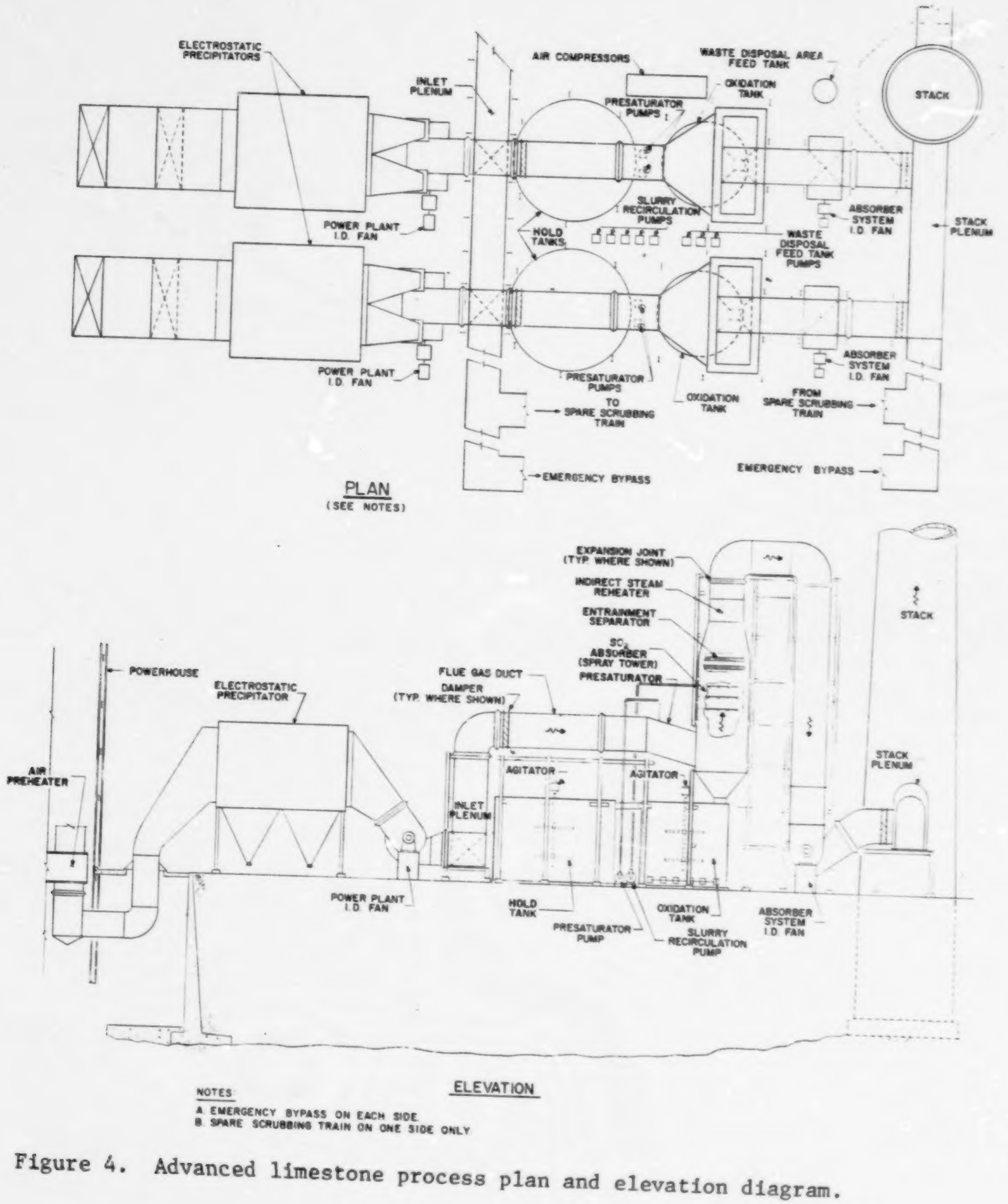

ELECTROSTATIC PRECIPITATORS
INLET PLENUM
AIR COMPRESSORS
WASTE DISPOSAL AREA FEED TANK
OXIDATION TANK
PRESATURATOR PUMPS
STACK
SLURRY RECIRCULATION PUMPS
ABSORBER SYSTEM I.D. FAN
STACK PLENUM
POWER PLANT I.D. FAN
HOLD TANKS
WASTE DISPOSAL FEED TANK PUMPS
POWER PLANT I.D. FAN
ABSORBER SYSTEM I.D. FAN
PRESATURATOR PUMPS
TO SPARE SCRUBBING TRAIN
OXIDATION TANK
FROM SPARE SCRUBBING TRAIN
EMERGENCY BYPASS
EMERGENCY BYPASS
PLAN
(SEE NOTES)
EXPANSION JOINT (TYP WHERE SHOWN)
INDIRECT STEAM REHEATER
ENTRAINMENT SEPARATOR
STACK
SO2 ABSORBER (SPRAY TOWER)
PRESATURATOR
FLUE GAS DUCT
DAMPER (TYP WHERE SHOWN)
POWERHOUSE
ELECTROSTATIC PRECIPITATOR
AIR PREHEATER
AGITATOR
AGITATOR
INLET PLENUM
STACK PLENUM
POWER PLANT I.D. FAN
HOLD TANK
PRESATURATOR PUMP
OXIDATION TANK
SLURRY RECIRCULATION PUMP
ABSORBER SYSTEM I.D. FAN
ELEVATION
NOTES
A. EMERGENCY BYPASS ON EACH SIDE
B. SPARE SCRUBBING TRAIN ON ONE SIDE ONLY

Figure 4. Advanced limestone process plan and elevation diagram.

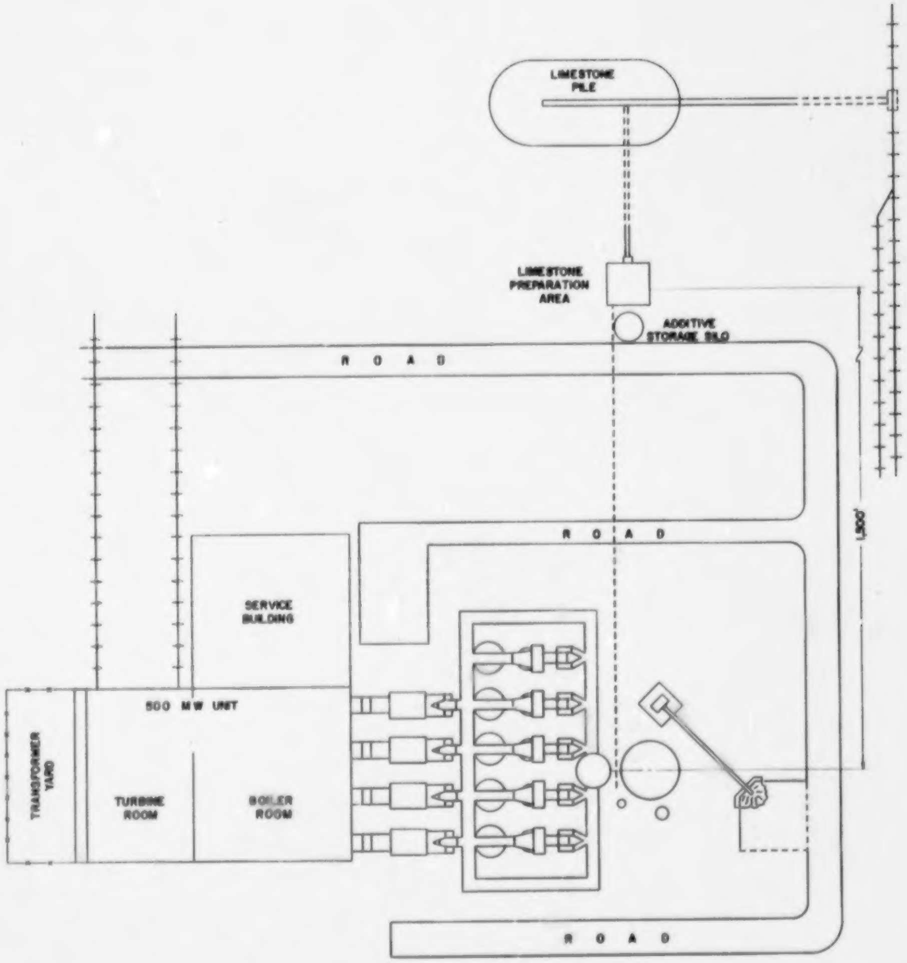

LIMESTONE PILE
LIMESTONE PREPARATION AREA
ADDITIVE STORAGE SILO
R O A D
1,300'
R O A D
SERVICE BUILDING
500 MW UNIT
TRANSFORMER YARD
TURBINE ROOM
BOILER ROOM
R O A D

Figure 5. Advanced limestone process plot plan.

TABLE 14. ADVANCED LIMESTONE PROCESS

EQUIPMENT LIST

Item (number): description	Total material cost, 1982 $
Area 1--Materials Handling	
1. Car shaker (1): Top mounting with crane, 20-hp shaker, 7-1/2-hp hoist	71,900
2. Car puller (1): 25-hp puller, 5-hp return	63,100
3. Hopper, limestone unloading (1): 16-ft diameter x 10-ft straight side, 2,570 ft3, carbon steel, 30-degree cone bottom	15,500
4. Feeder, limestone unloading (1): Vibrating pan, 3-1/2 hp, 250 tons/hr	5,500
5. Conveyor, limestone unloading (1): Belt, 36 in. wide x 20 ft long, 5 hp, 250 tons/hr, 165 ft/min	11,400
6. Conveyor, limestone unloading (incline) (1): Belt, 36 in. wide x 310 ft long, 50 hp, 15-degree incline, 80-ft total lift, 250 tons/hr, 165 ft/min	85,300
7. Dust collector, limestone unloading pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning, includes dust hood	11,200
8. Pump, limestone pit sump (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400
9. Conveyor, storage (1): Belt, 36 in. wide x 200 ft long, 5 hp, 250 tons/hr, 165 ft/min	73,100
10. Tripper, storage conveyor (1): 1 hp, 30 ft/min	27,200
11. Mobile equipment (1): Scraper tractor	141,900
12. Hopper, reclaim (2): 7-ft diameter x 4-1/4 ft deep x 2-ft bottom, 75 ft3, carbon steel, 60-degree cone bottom	2,400
13. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp, 100 tons/hr	10,900

(Continued)

TABLE 14. (Continued)

Item (number): description	Total material cost, 1982 $
14. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning	7,800
15. Pump, reclaim pit sump (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400
16. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 200 ft long, 5 hp, 100 tons/hr, 105 ft/min	40,900
17. Conveyor, limestone reclaim (incline) (1): Belt, 30 in. wide x 193 ft long, 40 hp, 15-degree incline, 50-ft total lift, 100 tons/hr, 105 ft/min	60,300
18. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90-ft lift, 100 tons/hr	57,800
19. Conveyor, live limestone feed (1): Belt, 30 in. wide x 60 ft long, 7-1/2 hp, 100 tons/hr, 105 ft/min	20,500
20. Tripper, live limestone feed conveyor (1): 1 hp, 30 ft/min	27,200
21. Bin, crusher feed (3): 13-ft diameter x 21-ft straight side height, 3,100 ft3, w/cover, 50-degree slope, carbon steel (2 operating, 1 spare)	43,300
Subtotal	782,000
Area 2--Feed Preparation	
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp (2 operating, 1 spare)	49,600
2. Crusher (3): Gyratory, 75 hp (2 operating, 1 spare)	297,100
3. Ball mill (3): Wet, open system, 9-ft diameter x 16 ft long, 575 hp, 250 tons/day (2 operating, 1 spare)	1,436,400
4. Dust collector, ball mill (3): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning (2 operating, 1 spare)	23,300

(Continued)

TABLE 14. (Continued)

Item (number): description	Total material cost, 1982 $
5. Tank, mills product (3): 10-ft diameter x 10 ft high, 5,500 gal, open top, four 10-in. baffles, agitator supports, carbon steel, flakeglass lined (2 operating, 1 spare)	13,700
6. Agitator, mills product tank (3): 40-in. diameter, 10 hp, neoprene coated (2 operating, 1 spare)	22,900
7. Pump, mills product tank (3): Centrifugal, 40 gpm, 60-ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,400
8. Silo, adipic acid storage (1): 8-1/2-ft diameter x 12-1/2-ft straight side height, 710 ft3, 60-degree slope, carbon steel	7,100
9. Conveyor, live adipic acid feed (1): Pneumatic, 10 hp	8,600
10. Bin, adipic acid feed (1): Carbon steel, rubber lined	2,100
11. Feeder, adipic acid addition (1): Screw, 6-in. diameter x 30 ft long, 1/2 hp, stainless steel	4,700
12. Dust collector, adipic acid feed bin (1): Bag filter, polypropylene bag, 450 aft3/min, 1-1/2 hp	3,300
13. Tank, slurry feed (1): 19-1/2-ft diameter x 19-1/2 ft high, 44,000 gal, open top, four 19-in. baffles, agitator supports, carbon steel, flakeglass lined	16,600
14. Agitator, slurry feed tank (1): 76-in. diameter, 40 hp, neoprene coated	35,500
15. Pump, slurry feed tank (8): Centrifugal, 21 gpm, 60-ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spares)	21,200
Subtotal	1,950,500

(Continued)

TYPING GUIDE SHEET

TABLE 14. (Continued)

Item (number): description	Total material cost, 1982 $
Area 3--Gas Handling	
1. **Fans** (5): Induced draft, 377,300 aft3/min, 7.1-in. static head, 600 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	3,122,400
Subtotal	3,122,400
Area 4--SO_2 Absorption	
1. **SO_2 absorber** (5): Spray tower, 40 ft high x 34 ft wide x 17 ft deep, 1/4-in. carbon steel, neoprene lined, 316 stainless steel grids, FRP chevron vane entrainment separator (4 operating, 1 spare)	5,081,300
2. **Tank, recirculation** (5): 34-ft diameter x 34 ft high, 227,700 gal, open top, four 34-in.-wide baffles, agitator supports, carbon steel, flakeglass lined (4 operating, 1 spare)	283,800
3. **Agitator, recirculation tank** (5): 136-in. diameter, 50 hp, neoprene coated (4 operating, 1 spare)	345,500
4. **Pump, presaturator** (10): Centrifugal, 1,294 gpm, 60-ft head, 57 hp, carbon steel, neoprene lined (4 operating, 6 spares)	101,000
5. **Pump, recirculation** (15): Centrifugal, 12,940 gpm, 100-ft head, 575 hp, carbon steel, neoprene lined (8 operating, 7 spares)	1,399,400
6. **Pump, makeup water** (2): Centrifugal, 3,230 gpm, 200-ft head, 275 hp, carbon steel (1 operating, 1 spare)	31,500
7. **Sootblowers** (40): Air, fixed	111,900
Subtotal	7,354,400
Area 5--Reheat	
1. **Reheaters** (5): Steam, tube type, 2,100 ft2, two-thirds of tubes made of Inconel 625 and one-third made of Cor-Ten (4 operating, 1 spare)	2,320,400

(Continued)

TABLE 14. (Continued)

Item (number): description	Total material cost, 1982 $
2. **Sootblowers** (20): Air, retractable	183,000
Subtotal	2,503,400
Area 6--Oxidation	
1. **Tank, oxidation** (5): 26-1/2-ft diameter x 34 ft high, 142,300 gal, open top, four 26-1/2-in.-wide baffles, agitator supports, carbon steel, flakeglass lined, includes air sparger (4 operating, 1 spare)	274,000
2. **Agitator, oxidation tank** (5): 102-in. diameter, 50 hp, neoprene coated (4 operating, 1 spare)	247,700
3. **Pump, oxidation tank** (8): Centrifugal, 378 gpm, 60-ft head, 10 hp, carbon steel, neoprene coated (4 operating, 4 spares)	36,700
4. **Blower, oxidation air** (6): 3,264 sft3/min, 300 hp (4 operating, 2 spares)	208,400
Subtotal	766,800
Area 7--Solids Separation	
1. **Tank, thickener feed** (1): 17-ft diameter x 34 ft high, 56,870 gal, open top, four 17-in.-wide baffles, agitator supports, carbon steel, flakeglass lined	22,600
2. **Agitator, thickener feed tank** (1): 68-in. diameter, 40 hp, neoprene coated	24,400
3. **Pump, thickener feed** (2): Centrifugal, 1,512 gpm, 60-ft head, 40 hp, carbon steel, neoprene lined (1 operating, 1 spare)	18,500
4. **Thickener** (1): 44-ft diameter x 5-1/2 ft high, carbon steel sides, concrete basin, includes 1-hp rake motor and mechanism	77,200
5. **Tank, thickener overflow** (1): 25-ft diameter x 5-1/2 ft high, 20,230 gal, open top, carbon steel	5,800

(Continued)

TABLE 14. (Continued)

Item (number): description	Total material cost, 1982 $
6. Pump, thickener overflow tank (2): Centrifugal, 1,226 gpm, 75-ft head, 40 hp, carbon steel, neoprene lined (1 operating, 1 spare)	12,200
7. Pump, thickener underflow (2): Centrifugal, 240 gpm, 10-ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,800
8. Tank, filter feed (1): 9-ft diameter x 9 ft high, 3,960 gal, open top, four 9-in. wide baffles, agitator supports, carbon steel, flakeglass lined	3,300
9. Agitator, filter feed tank (1): 36-in. diameter, 7-1/2 hp, neoprene coated	4,700
10. Pump, filter feed tank (3): Centrifugal, 120 gpm, 50-ft head, 3 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,800
11. Filter (3): Rotary vacuum, 8-ft diameter x 14-ft face, 50 total hp, includes auxiliary equipment (2 operating, 1 spare)	351,600
12. Pump, filtrate (4): Centrifugal, 84 gpm, 20-ft head, 1 hp, carbon steel, neoprene lined (2 operating, 2 spares)	17,200
13. Tank, filtrate surge (1): 7-1/2-ft diameter x 7-1/2 ft high, 2,780 gal, open top, carbon steel	1,500
14. Pump, filtrate surge tank (2): Centrifugal, 168 gpm, 85-ft head, 6 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,200
15. Conveyor, filter cake (1): Belt, 30 in. wide x 175 ft long, 1-1/2 hp, 50 tons/hr	37,100
16. Retaining wall, gypsum (1): 8 ft high x 32 ft long, concrete	1,600
Subtotal	606,500

Feed Preparation--

Limestone is reduced to about 0 x 3/4 inch using gyratory crushers and wet ground to 90% minus 325 mesh in two parallel ball mills. A third spare ball mill is provided. The 60% solids slurry is stored in the slurry feed tank (8-hour residence time), situated near the absorber system about 1,500 feet from the limestone preparation area. Adipic acid is pneumatically conveyed to a feed bin (8-hour storage) and metered into the mills product tanks by a screw feeder.

Gas Handling--

All flue gas ductwork between the power unit ID fans and the stack plenum is included in this area. A single inlet plenum receives the flue gas from the ID fans and distributes it to each of the five absorber trains (base case). An emergency bypass for 50% of the scrubbed gas is also included as two 25% capacity ducts, one around each end of the facility. An ID fan is provided in each absorber train to compensate for the 7.1-inch H_2O pressure drop (base case) through the FGD system. The clean gas is then ducted to the stack plenum.

SO_2 Absorption--

In each of the five trains, flue gas is cooled from 300°F to 124°F in a presaturator, using recycled slurry, before entering the spray tower absorber. The slurry is pumped from the recirculation tank to the absorber and sprayed through nozzles countercurrent to the gas flow. The absorber is operated at a stoichiometry of 1.07 mols of $CaCO_3$/mol (SO_2 + 2HCl) absorbed with a L/G ratio of 80 gal/kaft^3. The absorber contains four banks of slurry spray nozzles. A chevron vane entrainment separator is used to reduce liquid carryover; it is equipped with top and bottom wash using fresh makeup water.

The absorber effluent drains into an oxidation tank beneath the absorber (the cost of this tank is included in the oxidation area costs), from which it is pumped to an agitated recirculation tank with an 8-minute residence time. Makeup slurry is added to this tank. The regenerated slurry is pumped from the recirculation tank to the absorber spray headers with two pumps. A spare pump is provided.

Reheat--

Each train contains an indirect steam reheater equipped with sootblowers for cleaning. The reheater is designed to achieve a flue gas temperature of 175°F at the stack to evaporate entrained moisture and increase plume buoyancy. The effects of the heat of compression in the FGD ID fan and heat from the bypassed flue gas are included in the design.

Oxidation--

Absorber effluent drains into an oxidation tank beneath each absorber. The tank has a 5-minute residence time and is equipped with an agitator and an air sparging manifold. A blower forces air through spargers at a stoichiometry of 2.5 atoms O/mol SO_2 absorbed to oxidize the sulfites to gypsum. A bleedstream is pumped to the solids separation area. An overflow stream flows by gravity to the recirculation tank for reuse.

Solids Separation--

The oxidation tank bleedstream is concentrated to 40% solids in a thickener and further dewatered to 85% solids in a rotary vacuum filter. The filtrate is combined with the thickener overflow and returned to the oxidation tank and the wet ball mills. The dewatered gypsum is conveyed to an in-process storage area and is loaded into dump trucks by a wheeled loader for transport to a landfill situated one mile from the FGD facility. The landfill is described in the premises. For the base case, the landfill is square with sides of approximately 1,800 feet and occupies 106 acres.

DAVY S-H PROCESS

The Davy S-H process design used in this study represents the state of technology of the Saarberg-Holter process in 1980 as supplied by Davy Corporation, Ltd. In some cases, the recommended design is modified to be compatible with the other processes. These modifications are designated in the following discussion.

The common plenum distributes the flue gas to the Rotopart absorbers, a patented modular design spray column. SO_2 is removed from the flue gas by cocurrent scrubbing with a clear alkaline solution. Absorber effluent is oxidized with air, mixed with lime to regenerate the scrubbing solution, and the resulting waste is dewatered in a thickener and rotary vacuum filters. The gypsum cake is transported to a landfill for disposal. The Davy S-H process flow diagram is shown in Figure 6 and the base case material balance is shown in Table 15. The process design drawings for the base case are shown in Figures 7 and 8. The equipment list, itemized by area, and the cost of the equipment are shown in Table 16. The costs do not include installation.

Major Processing Areas

The process is divided into seven processing areas: materials handling, feed preparation, gas handling, SO_2 absorption, stack gas reheat, oxidation, and solids separation and disposal.

Materials Handling--

Pebble lime, delivered by rail, is transported by an underground conveyor and stored in two 15-day-capacity silos. Underground conveyors also transport lime to two feed bins, each with a 12-hour capacity. The formic acid and ferric chloride addition systems, 30-day-capacity storage tanks, and metering pumps are also included in this area.

Feed Preparation--

Two duplicate slaking systems are used to produce the lime slurry. Lime from the feed bin is metered to the slaker by screw feeders where enough fresh water is added to produce a 15% slurry after reaction. The slaker overflows into a slurry buffer tank with a 10-minute residence time and is then pumped

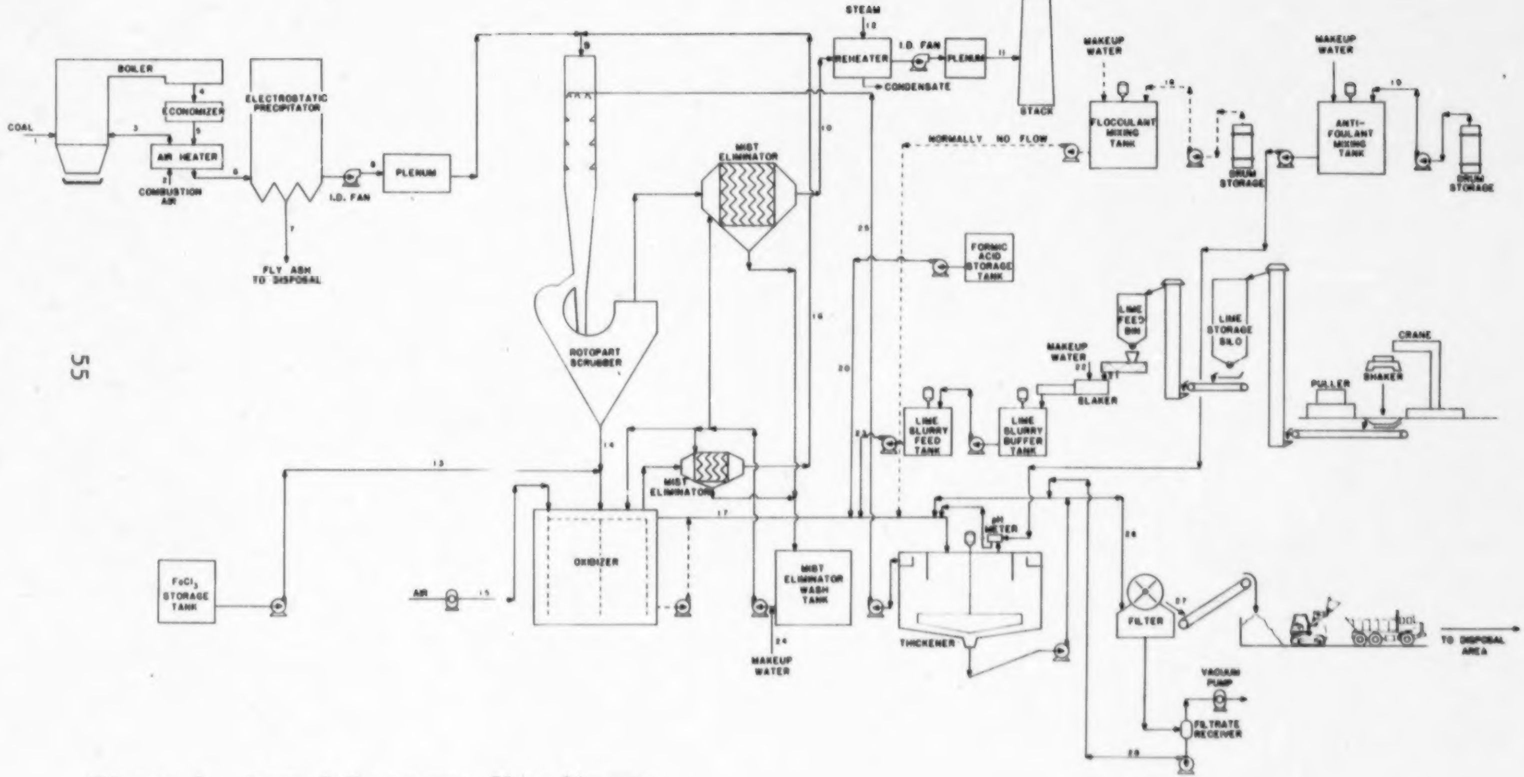

COAL
1
BOILER
4
ECONOMIZER
3
5
AIR HEATER
2
COMBUSTION AIR
6
ELECTROSTATIC PRECIPITATOR
7
FLY ASH TO DISPOSAL
I.D. FAN
8
PLENUM
9
ROTOPART SCRUBBER
MIST ELIMINATOR
10
16
14
13
MIST ELIMINATOR
17
FeCl$_3$ STORAGE TANK
AIR
15
OXIDIZER
MIST ELIMINATOR WASH TANK
24
MAKEUP WATER
STEAM
12
REHEATER
I.D. FAN
PLENUM
11
CONDENSATE
STACK
25
20
23
NORMALLY NO FLOW
FORMIC ACID STORAGE TANK
MAKEUP WATER
FLOCCULANT MIXING TANK
18
DRUM STORAGE
MAKEUP WATER
ANTI-FOULANT MIXING TANK
19
DRUM STORAGE
LINE SLURRY FEED TANK
LINE SLURRY BUFFER TANK
MAKEUP WATER
22
SLAKER
LIME FEED BIN
LIME STORAGE SILO
PULLER
SHAKER
CRANE
pH METER
THICKENER
26
FILTER
27
TO DISPOSAL AREA
VACUUM PUMP
FILTRATE RECEIVER
28

Figure 6. Davy S-H process flow diagram.

TABLE 15. DAVY S-H PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Total air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,072,000	4,378,000	4,772,000	4,772,000
2						
3	Flow rate, sft^3/min(60°F)		1,120,000	967,300	1,003,000	1,003,000
4	Temperature, °F		80	495		705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Fly ash to disposal	Gas to plenum	Gas to Rotopart	Gas to reheater
1	Total stream, lb/hr	5,465,000	50,050	5,415,000	5,533,000	5,656,000
2						
3	Flow rate, sft^3/min(60°F)	1,156,000		1,157,000	1,183,000	1,230,000
4	Temperature, °F	300		300	296	115
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Gas to stack	Steam to reheater	Ferric chloride to oxidizer	Scrubber effluent to oxidizer	Air to oxidizer
1	Total stream, lb/hr	5,656,000	121,600	159	25,110,000	117,600
2						
3	Flow rate, sft^3/min(60°F)	1,232,000				25,980
4	Temperature, °F	175	470			200
5	Pressure, psig		500			13
6	Flow rate, gpm			0.2	49,710	
7	Specific gravity			1.32	1.01	
8	pH				4.0-4.5	
9	Undissolved solids, %					
10						

	Stream No.	16	17	18	19	20
	Description	Oxidizer off-gas to Rotopart	Oxidizer overflow to thickener	Flocculant to mixing tank	Antifoulant to mixing tank	Formic acid to thickener
1	Total stream, lb/hr	118,300	25,120,000	(Normally no flow)	0.3	81
2						
3	Flow rate, sft^3/min(60°F)	27,070				
4	Temperature, °F	115				
5	Pressure, psig	0.1				
6	Flow rate, gpm		49,740		-	0.1
7	Specific gravity		1.01	1.03	1.15	1.21
8	pH		3.5-4.0			
9	Undissolved solids, %					
10						

(Continued)

TABLE 15. (Continued)

Stream No.		21	22	23	24	25
Description		Lime to slaker	Makeup water to slaker	Lime slurry to thickener	Makeup water to mist eliminator wash tank	Thickener overflow to Rotopart
1	Total stream, lb/hr	21,080	162,300	183,400	12,100	25,227,000
2						
3	Flow rate, sft^3/min(60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm		325	340	24	49,975
7	Specific gravity		1.0	1.08	1.0	1.01
8	pH					
9	Undissolved solids, %			15		
10						

Stream No.		26	27	28		
Description		Gypsum slurry to filter	Gypsum cake to disposal	Filtrate to thickener		
1	Total stream, lb/hr	407,400	71,910	335,500		
2						
3	Flow rate, sft^3/min(60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm	741		658		
7	Specific gravity	1.1	2.0	1.02		
8	pH					
9	Undissolved solids, %	15	85			
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

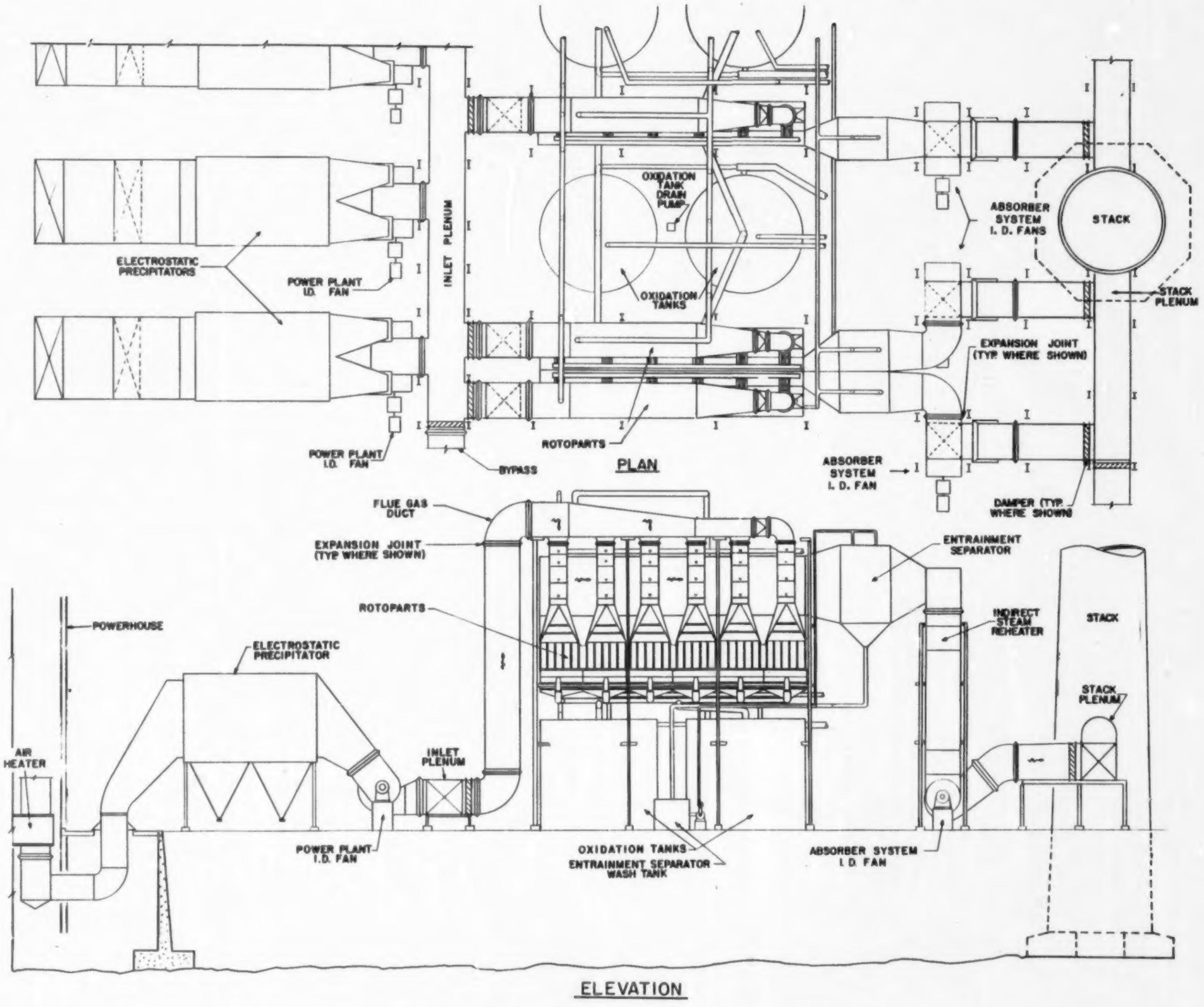

ELECTROSTATIC PRECIPITATORS
POWER PLANT I.D. FAN
POWER PLANT I.D. FAN
INLET PLENUM
BYPASS
OXIDATION TANK DRAIN PUMP
OXIDATION TANKS
ROTOPARTS
PLAN
ABSORBER SYSTEM I. D. FANS
STACK
STACK PLENUM
EXPANSION JOINT (TYP. WHERE SHOWN)
ABSORBER SYSTEM I. D. FAN
DAMPER (TYP. WHERE SHOWN)
FLUE GAS DUCT
EXPANSION JOINT (TYP WHERE SHOWN)
ROTOPARTS
POWERHOUSE
ELECTROSTATIC PRECIPITATOR
AIR HEATER
INLET PLENUM
POWER PLANT I.D. FAN
ENTRAINMENT SEPARATOR
INDIRECT STEAM REHEATER
STACK
STACK PLENUM
OXIDATION TANKS
ENTRAINMENT SEPARATOR WASH TANK
ABSORBER SYSTEM I.D. FAN
ELEVATION

Figure 7. Davy S-H process plan and elevation diagram.

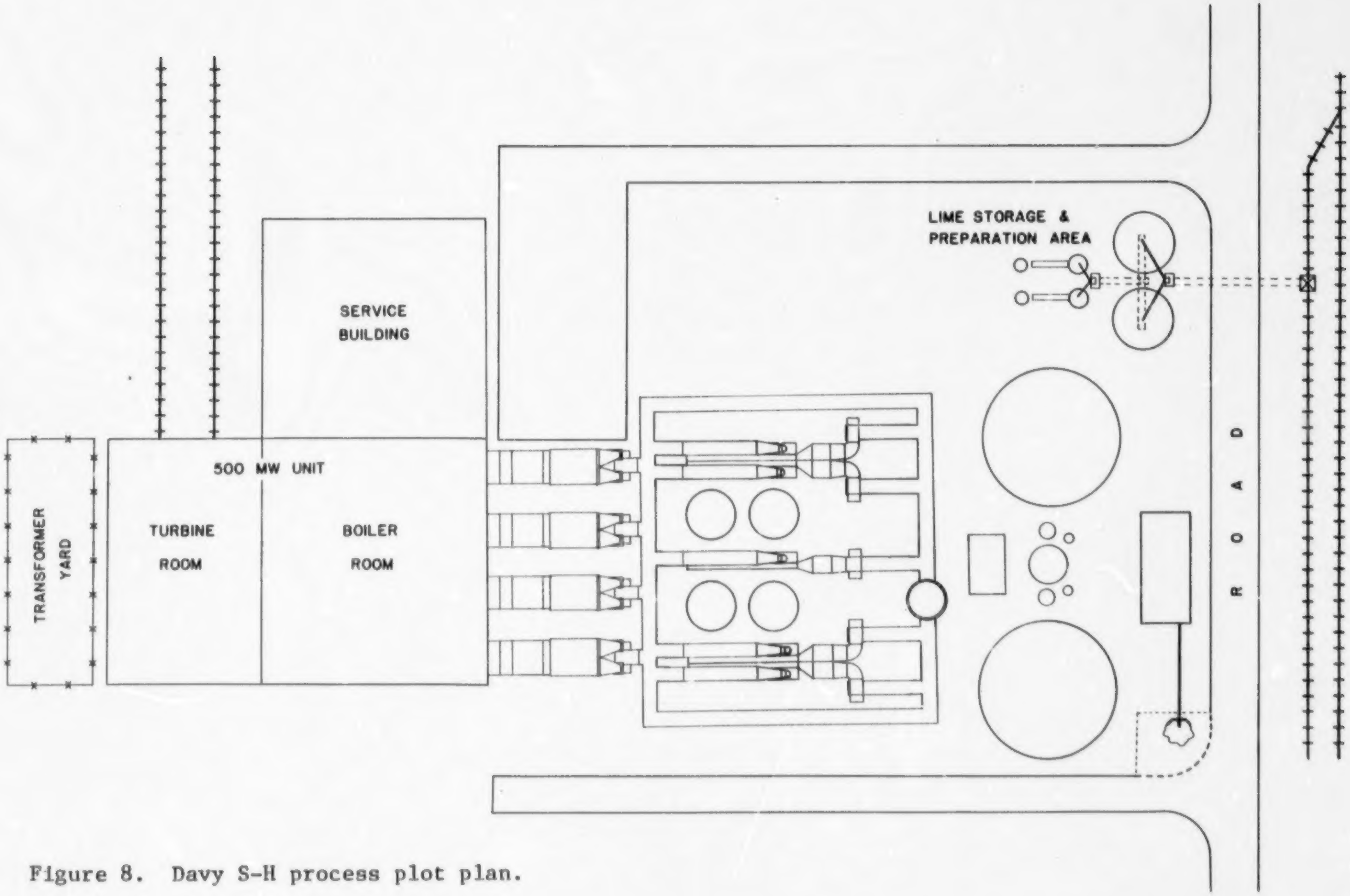

LIME STORAGE &
PREPARATION AREA
SERVICE
BUILDING
500 MW UNIT
TURBINE
ROOM
BOILER
ROOM
TRANSFORMER
YARD
R O A D

Figure 8. Davy S-H process plot plan.

TABLE 16. DAVY S-H PROCESS

EQUIPMENT LIST

Item (number): description	Total material cost, 1982 $
Area 1--Materials Handling	
1. Car shaker (1): Top mounting with crane, 20-hp shaker, 7-1/2-hp hoist	71,900
2. Car puller (1): 25-hp puller, 5-hp return	63,100
3. Hopper, unloading (1): 12 ft x 12 ft x 2-ft bottom, 26 ft deep, 1,500 ft3, carbon steel	7,400
4. Pump, pit sump (3): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined (2 operating, 1 spare)	7,200
5. Conveyor, lime unloading (enclosed) (1): Belt, 24 in. wide x 102 ft long, 2 hp, 100 tons/hr, 180 ft/min	19,300
6. Elevator, lime storage (1): Continuous bucket, 20 in. x 8 in. x 11-3/4 in., 25 hp, 112-ft lift, 100 tons/hr, with diverter gate	42,700
7. Dust collector, unloading pit (1): Bag filter, polyester bag, 2,200 aft3/min, 7-1/2 hp, air-pulse cleaning	7,600
8. Silo, lime storage (2): 49-ft diameter x 72-ft straight side height, 135,800 ft3, 45-degree slope, concrete	370,000
9. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp, 40 tons/hr	10,900
10. Conveyor, lime silo unloading (enclosed) (2): Belt, 18 in. wide x 35 ft long, 1 hp, 15 tons/hr, 50 ft/min	17,900
11. Conveyor, lime reclaim (enclosed) (1): Belt, 18 in. wide x 41 ft long, 1 hp, 30 tons/hr, 100 ft/min	9,900
12. Elevator, live lime feed (1): Continuous bucket, 12 in. x 6 in. x 11-3/4 in., 5 hp, 69-ft lift, 35 tons/hr, with diverter gate	18,300

(Continued)

TYPING GUIDE SHEET

TABLE 16. (Continued)

Item (number): description	Total material cost, 1982 $
13. Bin, lime feed (2): 16-ft diameter x 23-ft straight side height, 4,600 ft3, w/cover, 45-degree slope, carbon steel (1 operating, 1 spare)	32,300
14. Dust collecting system, slaker feeder (1): Bag filter, polyester bag, 8,800 aft3/min, 25 hp, air-pulse cleaning (1/2 cost in feed preparation area)	8,300
15. Tank, ferric chloride storage (1): 12-1/2-ft diameter x 12-1/2 ft high, 11,500 gal, w/cover, carbon steel, neoprene lined	11,500
16. Pump, ferric chloride metering (2): Diaphragm, 15 gph, 35-ft head, 1/2 hp, filled Teflon (1 operating, 1 spare)	3,300
17. Tank, formic acid storage (1): 10-ft diameter x 10 ft high, 5,900 gal, domed top, fiberglass	5,700
18. Pump, formic acid metering (2): Duplex, 40 gph, 50-ft head, 1/2 hp, filled Teflon (1 operating, 1 spare)	4,000
Subtotal	711,300
Area 2--Feed Preparation	
1. Feeder, lime feed bin discharge (2): Vibrating, 3-1/2 hp (1 operating, 1 spare)	10,900
2. Feeder, slaker (2): Screw, 12-in. diameter x 12 ft long, 1 hp (1 operating, 1 spare)	11,000
3. Slaker (2): 7 ft wide x 31 ft long, 10 hp, 2-hp classifier, 10 tons/hr (1 operating, 1 spare)	60,000
4. Tank, slurry buffer (2): 10-ft diameter x 6 ft high, 3,500 gal, open top, four 10-in. baffles, agitator supports, carbon steel, neoprene lined (1 operating, 1 spare)	6,900
5. Agitator, slurry buffer tank (2): 40-in. diameter, 2-1/2 hp, neoprene coated (1 operating, 1 spare)	5,900
6. Pump, slurry buffer tank (3): Centrifugal, 170 gpm, 60-ft head, 7-1/2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	13,800

(Continued)

TABLE 16. (Continued)

Item (number): description	Total material cost, 1982 $
7. Tank, slurry feed (1): 30-ft diameter x 31 ft high, 163,900 gal, open top, four 30-in. baffles, agitator supports, carbon steel, neoprene lined	64,600
8. Agitator, slurry feed tank (1): 120-in. diameter, 50 hp, neoprene coated	59,900
9. Pump, slurry feed tank (2): Centrifugal, 340 gpm, 60-ft head, 15 hp, carbon steel, neoprene lined (1 operating, 1 spare)	10,400
10. Dust collecting system, slaker feeder (1): Bag filter, polyester bag, 8,800 aft3/min, 25 hp, air-pulse cleaning (1/2 cost in materials handling area)	8,300
11. Agitator, flocculant drum (1): 6-in. diameter, 1/2 hp, 316 stainless steel	700
12. Pump, flocculant transfer (2): Diaphragm, 30 gph, 60-ft head, 1/2 hp, filled Teflon (1 operating, 1 spare)	3,700
13. Tank, flocculant mixing (1): 6-ft diameter x 6 ft high, 1,270 gal, open top, four 6-in. baffles, agitator supports, carbon steel	1,100
14. Agitator, flocculant mixing tank (1): 24-in. diameter, 2 hp	2,500
15. Pump, flocculant feed (2): Gear, 20 gpm, 60-ft head, 1/2 hp, filled Teflon (1 operating, 1 spare)	6,000
16. Pump, antifoulant transfer (2): Diaphragm, 5 gpm, 10-ft head, 1/3 hp, 316 stainless steel casing, SS Vitron TFE Buna N diaphragm (1 operating, 1 spare)	3,000
17. Tank, antifoulant mixing (1): 8-ft diameter x 8 ft high, 3,000 gal, open top, four 8-in. baffles, agitator supports, carbon steel, phenolic resin lined	2,500
18. Agitator, antifoulant mixing tank (1): 32-in. diameter, 3-1/2 hp, 316 stainless steel	6,200

(Continued)

TABLE 16. (Continued)

Item (number): description	Total material cost, 1982 $
19. Pump, antifoulant metering (2): Diaphragm, 90 gph, 50-ft head, 1/2 hp, 316 stainless steel casing, SS Vitron TFE Buna N diaphragm (1 operating, 1 spare)	4,800
Subtotal	282,200
Area 3--Gas Handling	
1. Fans (5): Induced draft, 376,100 aft3/min, 15-in. static head, 1,500 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	4,488,900
Subtotal	4,488,900
Area 4--SO_2 Absorption	
1. SO_2 absorber (15): Rotopart, 30 ft x 15 ft x 52 ft high, 2 washing ducts (6-ft diameter), neoprene lined, carbon steel, plastic chevron vane entrainment separator (12 operating, 3 spares)	5,582,000
2. Pump, absorber feed (15): Centrifugal, 6,250 gpm, 160-ft head, 450 hp, carbon steel, neoprene lined (12 operating, 3 spares)	1,020,100
3. Pump, mist eliminator wash (4): Centrifugal, 375 gpm, 125-ft head, 25 hp, carbon steel, neoprene lined (2 operating, 2 spares)	23,500
4. Tank, mist eliminator wash (2): 11-ft diameter x 10 ft high, 7,100 gal, open top, carbon steel, neoprene lined	13,700
5. Pump, makeup water (2): Centrifugal, 364 gpm, 75-ft head, 15 hp, carbon steel (1 operating, 1 spare)	9,800
6. Sootblowers (40): Air, fixed	111,900
Subtotal	6,761,000

(Continued)

TYPING GUIDE SHEET

TABLE 16. (Continued)

Item (number): description	Total material cost, 1982 $
Area 5--Reheat	
1. Reheater (5): Steam, tube type, 3,400 ft2, one-half of tubes made of Inconel 625 and one-half made of Cor-Ten (4 operating, 1 spare)	2,974,800
2. Sootblowers (20): Air, retractable	183,000
Subtotal	3,157,800
Area 6--Oxidation	
1. Tank, oxidation (4): 40-ft diameter x 43 ft high, 404,200 gal, w/cover, carbon steel, neoprene lined, includes 316 stainless steel sparger and downcomer (4 operating, no spare)	653,500
2. Blower, oxidation air (6): 14 psig, 6,610 aft3/min, 500 hp (4 operating, 2 spares)	327,100
3. Pump, oxidizer drain (2): Centrifugal, 650 gpm, 60-ft head, 20 hp, carbon steel, neoprene lined (2 spares)	15,400
4. Mist eliminator, oxidizer off-gas (4): Chevron vane, 30,000 aft3/min, plastic blades, fiberglass frame	16,000
Subtotal	1,012,000
Area 7--Solids Separation	
1. Thickener (2): 114-ft diameter x 28 ft high, carbon steel sides, concrete basin, includes 5-hp rake motor and mechanism	492,700
2. Pump, thickener underflow (4): Centrifugal, 370 gpm, 50-ft head, 10 hp, carbon steel, neoprene lined (2 operating, 2 spares)	20,100
3. Tank, filter feed (1): 13-ft diameter x 12-1/2 ft high, 12,400 gal, open top, four 13-in.-wide baffles, agitator supports, carbon steel, neoprene lined	11,900

(Continued)

TABLE 16. (Continued)

Item (number): description	Total material cost, 1982 $
4. Agitator, filter feed tank (1): 52-in. diameter, 10 hp, neoprene lined	9,700
5. Pump, filter feed tank (8): Centrifugal, 185 gpm, 50-ft head, 7-1/2 hp, carbon steel, neoprene lined (4 operating, 4 spares)	33,400
6. Filter (3): Rotary vacuum, 10-ft diameter x 10-ft face, 50 total hp, includes auxiliary equipment (2 operating, 1 spare)	345,600
7. Pump, filtrate (2): Centrifugal, 643 gpm, 70-ft head, 25 hp, carbon steel, neoprene lined (1 operating, 1 spare)	16,000
8. Tank, filtrate surge (1): 12-ft diameter x 12-1/2 ft high, 10,600 gal, open top, carbon steel	4,400
9. Pump, filtrate surge tank (2): Centrifugal, 643 gpm, 70-ft head, 25 hp, carbon steel, neoprene lined (1 operating, 1 spare)	16,000
10. Conveyor, filter cake (1): Belt, 30 in. wide x 180 ft long, 1-1/2 hp, 40 tons/hr	38,000
11. Retaining wall, gypsum (1): 8 ft high x 32 ft long, concrete	1,600
Subtotal	989,400

into a single slurry feed tank with an 8-hour residence time. The slurry is added to the effluent from the oxidation tanks.

The antifoulant and flocculent addition systems are included in this area. Antifoulant is used in the thickener pH measuring system to prevent calcium buildup in the instrument. Flocculent is added, if required because of an ESP upset, to aid in settling of fly ash.

Gas Handling--

This area includes all flue gas ductwork between the power unit ID fans and the stack plenum. An inlet plenum receives the flue gas from the power unit ID fans and distributes it to the five absorber trains (base case). Flue gas leaves the plenum at 300°F and is mixed with the oxidizer off-gas at 115°F, producing a gaseous mixture at 296°F. An ID fan downstream of the reheater is provided in each train to compensate for the 15-inch H_2O pressure drop in the FGD system. [Davy recommends a forced-draft (FD) fan for all installations.] The flue gas is then ducted to the stack plenum. An emergency bypass duct is included in this area to handle up to 50% of the flue gas. The bypass is installed as two 25% capacity ducts, one around each end of the FGD facility.

SO_2 Absorption--

The Davy S-H SO_2 absorber or Rotopart is a patented modular design cocurrent spray column. Each module consists of two absorption tubes and a single separator section. The vertical absorption tubes contain water shedding rings (venturi throats) to separate the inactive fluid from the tube walls and recontact it with the flue gas. Several modules can be connected by a single header to form an absorption train. For the base case, five trains of three modules each are used of which four trains are operational and one train is a spare. (Davy does not recommend a spare train.) From the absorption tubes, the flue gas and absorbent liquid flow into the separator section where the water and gas are separated in three steps. First, a sudden change in flow direction removes the larger droplets that constitute most of the liquid. Next, the gas is forced into a circular path in which centrifugal force removes more droplets. Finally, the gas enters a chevron vane mist eliminator where most of the remaining liquid is removed.

Reheat--

Equipment in this area includes an indirect steam reheater and sootblowers for each train. (Davy states that sootblowers are unnecessary.) The reheater increases the temperature of the cleaned gas to 170°F to evaporate entrained moisture and increase plume buoyancy. The flue gas then passes through the ID fan where heat of compression increases the temperature to 175°F and enters the stack plenum.

Oxidation--

The acidic effluent from each absorber train flows to an oxidation tank with a 30-minute residence time for oxidation of the sulfite species to sulfate. The four oxidizers are arranged so that each operating train uses one oxidizer. A blower serves each oxidizer to supply reaction air and agitate the mixture to keep the gypsum crystals in suspension. The oxidized slurry overflows into a mixing channel for transport to the thickeners. Oxidizer off-gas (excess air) is returned to the flue gas entering the Rotopart.

Solids Separation--

Two 50% capacity thickeners (90-minute residence time) concentrate the oxidizer overflow to 15% solids. This stream is further dewatered using rotary vacuum filters to produce a cake of 85% solids. The filtrate is pumped back to the thickener. The dewatered gypsum is conveyed to an in-process storage pile where wheeled loaders transfer the gypsum to dump trucks for transport to a landfill situated one mile from the FGD facility. The landfill used for the base case, as described in the premises, is approximately 1,760 feet on a side and covers a total of 103 acres.

DOWA PROCESS

The design of the Dowa process used in this study represents the state of technology of the Dowa process in 1980 as supplied by UOP, Inc., the U.S. licensee, based in part on preliminary Dowa process demonstration data at the Shawnee test facility.

A solution of basic aluminum sulfate absorbs SO_2 in a countercurrent packed tower and is oxidized by air to aluminum sulfate complexes. The oxidized liquid is regenerated to basic aluminum sulfate by neutralization with limestone. The gypsum byproduct from the neutralization step is concentrated in a thickener, dewatered by filters, and disposed of in a landfill. A flow diagram and material balance are shown in Figure 9 and Table 17. The Dowa process design drawings for the base case are shown in Figures 10 and 11. The equipment list, itemized by area, and the purchase cost of the equipment are shown in Table 18. The costs do not include installation.

Major Processing Areas

The Dowa process is divided into nine processing areas: materials handling, feed preparation, gas handling, SO_2 absorption, stack gas reheat, oxidation, neutralization, chloride purge, and solids separation and disposal.

Materials Handling--

Limestone, crushed and screened to 0 x 1-1/2 inches at the quarry and containing 95% $CaCO_3$, is delivered by truck or rail and unloaded by conveyor to a 30-day stockpile. Facilities also include conveyors and elevators to transport limestone approximately 150 feet from the stockpile to the limestone preparation area. Aluminum sulfate is pneumatically conveyed from truck or railcar to a 30-day silo.

Feed Preparation--

Limestone is reduced to about 0 x 3/4 inch using gyratory crushers and wet ground to 90% minus 325 mesh in two parallel operating ball mills. A

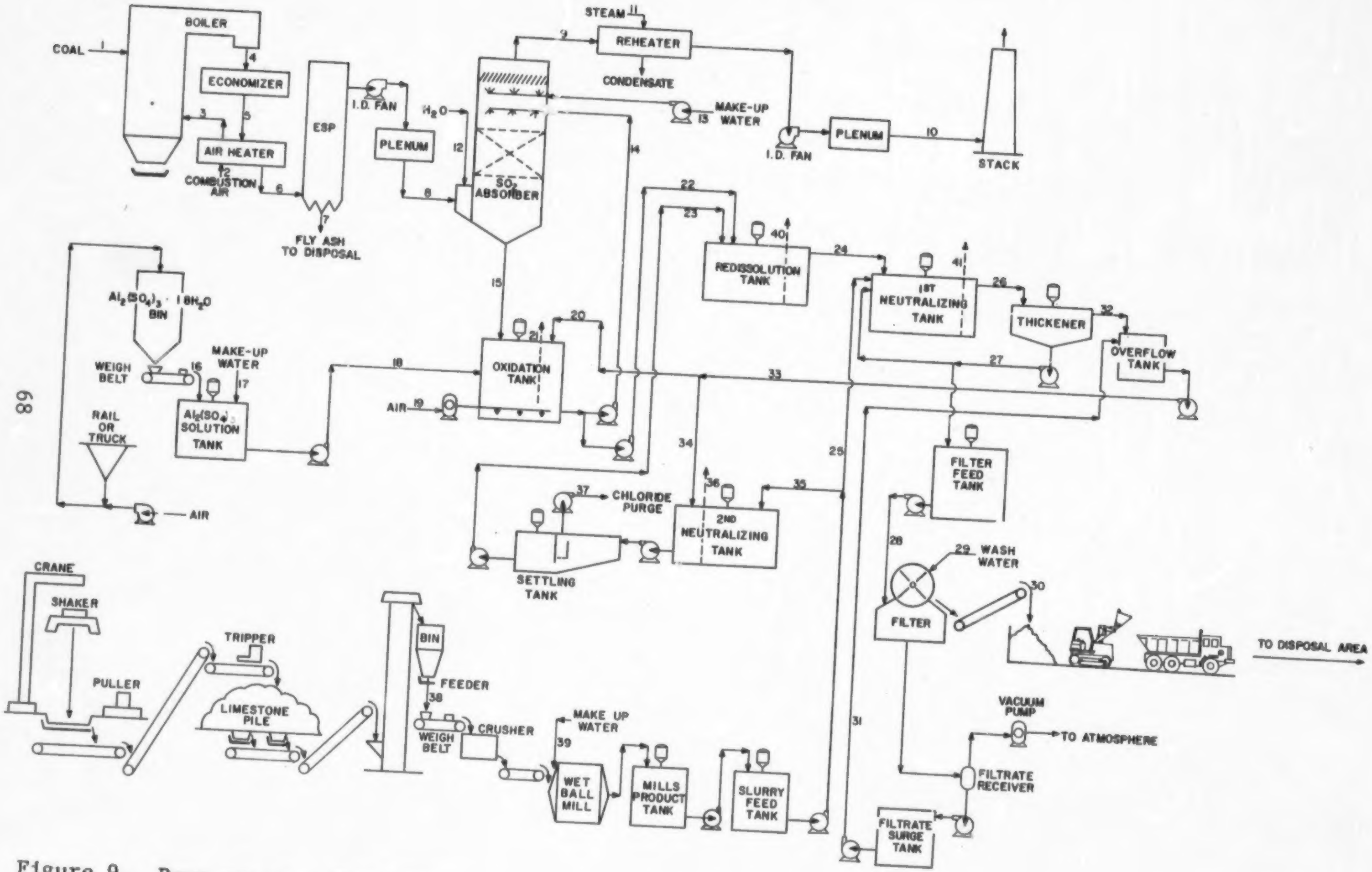

COAL
1
BOILER
4
ECONOMIZER
3
5
AIR HEATER
12
COMBUSTION AIR
6
ESP
7
FLY ASH TO DISPOSAL
I.D. FAN
PLENUM
8
H_2O
12
SO_2 ABSORBER
9
STEAM 11
REHEATER
CONDENSATE
13 MAKE-UP WATER
14
I.D. FAN
PLENUM
10
STACK
15
$Al_2(SO_4)_3 \cdot 18H_2O$ BIN
WEIGH BELT
16
MAKE-UP WATER
17
$Al_2(SO_4)_3$ SOLUTION TANK
RAIL OR TRUCK
AIR
18
AIR 19
OXIDATION TANK
21
20
22
23
REDISSOLUTION TANK
40
24
1ST NEUTRALIZING TANK
41
26
THICKENER
32
OVERFLOW TANK
27
33
34
25
35
36
2ND NEUTRALIZING TANK
37 CHLORIDE PURGE
SETTLING TANK
FILTER FEED TANK
28
29 WASH WATER
30
FILTER
TO DISPOSAL AREA
VACUUM PUMP
TO ATMOSPHERE
FILTRATE RECEIVER
FILTRATE SURGE TANK
31
CRANE
SHAKER
PULLER
TRIPPER
LIMESTONE PILE
BIN
FEEDER
38
WEIGH BELT
CRUSHER
MAKE UP WATER
39
WET BALL MILL
MILLS PRODUCT TANK
SLURRY FEED TANK

Figure 9. Dowa process flow diagram.

TABLE 17. DOWA PROCESS

MATERIAL BALANCE

Stream No.		1	2	3	4	5
Description		**Coal to boiler**	**Total air to air heater**	**Combustion air to boiler**	**Gas to economizer**	**Gas to air heater**
1	Total stream, lb/hr	406,000	5,072,000	4,378,000	4,772,000	4,772,000
2						
3	Flow rate, sft³/min(60°F)		1,120,000	967,300	1,003,000	1,003,000
4	Temperature, °F		80	495		705
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	Basicity					
9	Undissolved solids, %					
10						

Stream No.		6	7	8	9	10
Description		**Gas to electrostatic precipitator**	**Fly ash to disposal**	**Gas to absorber**	**Gas to reheater**	**Gas to stack**
1	Total stream, lb/hr	5,465,000	50,050	5,415,000	5,641,000	5,641,000
2						
3	Flow rate, sft³/min(60°F)	1,156,000		1,157,000	1,240,000	1,242,000
4	Temperature, °F	300		300	125	175
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	Basicity					
9	Undissolved solids, %					
10						

Stream No.		11	12	13	14	15
Description		**Steam to reheater**	**Water to wall wash**	**Water to mist eliminator wash**	**Scrubbing liquor to absorber**	**Absorber effluent to oxidation tank**
1	Total stream, lb/hr	104,900	60,030	182,500	35,294,000	35,310,000
2						
3	Flow rate, sft³/min(60°F)					
4	Temperature, °F	470				
5	Pressure, psig	500				
6	Flow rate, gpm		120	373	58,850	58,870
7	Specific gravity				1.20	1.20
8	Basicity				10.0	9.0
9	Undissolved solids, %					
10						

Stream No.		16	17	18	19	20
Description		**Makeup aluminum sulfate**	**Water to makeup solution tank**	**Makeup aluminum sulfate solution**	**Oxidation air**	**Regenerated effluent to oxidation tank**
1	Total stream, lb/hr	506	4,496	5,002	95,570	2,311,000
2						
3	Flow rate, sft³/min(60°F)				21,120	
4	Temperature, °F				80	
5	Pressure, psig					
6	Flow rate, gpm		8	9		3,886
7	Specific gravity			1.22		1.19
8	Basicity					25.0
9	Undissolved solids, %					
10						

(Continued)

TABLE 17. (Continued)

	Stream No.	21	22	23	24	25
	Description	Oxidation tank off-gas	Oxidized effluent to redissolution tank	Regenerated effluent from settling tank	Liquor from redissolution tank	Limestone slurry to 1st neutralizing tank
1	Total stream, lb/hr	95,970	2,334,000	132,000	2,465,000	38,520
2						
3	Flow rate, sft^3/min(60°F)	22,130				
4	Temperature, °F	120				
5	Pressure, psig					
6	Flow rate, gpm		3,888	230	4,110	48
7	Specific gravity		1.20	1.15	1.20	1.6
8	Basicity		10.0		16.0	
9	Undissolved solids, %			20	1	60
10						

	Stream No.	26	27	28	29	30
	Description	Slurry to thickener	Thickener underflow	Thickener underflow to filter	Filter wash water	Gypsum cake to disposal
1	Total stream, lb/hr	3,493,000	1,175,000	175,900	37,630	72,830
2						
3	Flow rate, sft^3/min(60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm	5,542	1,621	243	75	
7	Specific gravity	1.26	1.45	1.45		
8	Basicity	25.0	25.0	25.0		
9	Undissolved solids, %	12	35	35		85
10						

	Stream No.	31	32	33	34	35
	Description	Filtrate return	Thickener overflow	Regenerated effluent from overflow tank	Regenerated effluent to 2nd neutralizing tank	Limestone slurry to 2nd neutralizing tank
1	Total stream, lb/hr	140,700	2,317,000	2,458,000	147,100	22,790
2						
3	Flow rate, sft^3/min(60°F)					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm	245	3,893	4,132	248	29
7	Specific gravity	1.15	1.19	1.19	1.19	1.6
8	Basicity	25.0	25.0	25.0	25.0	
9	Undissolved solids, %					60
10						

	Stream No.	36	37	38	39	40
	Description	CO_2 vent from 2nd neutralizing tank	Chloride purge	Limestone to weigh feeder	Water to wet ball mill	CO_2 vent from redissolution tank
1	Total stream, lb/hr	4,775	32,800	38,950	22,360	1,792
2						
3	Flow rate, sft^3/min(60°F)	686				
4	Temperature, °F					258
5	Pressure, psig					
6	Flow rate, gpm		64		45	
7	Specific gravity		1.02			
8	Basicity					
9	Undissolved solids, %					
10						

(Continued)

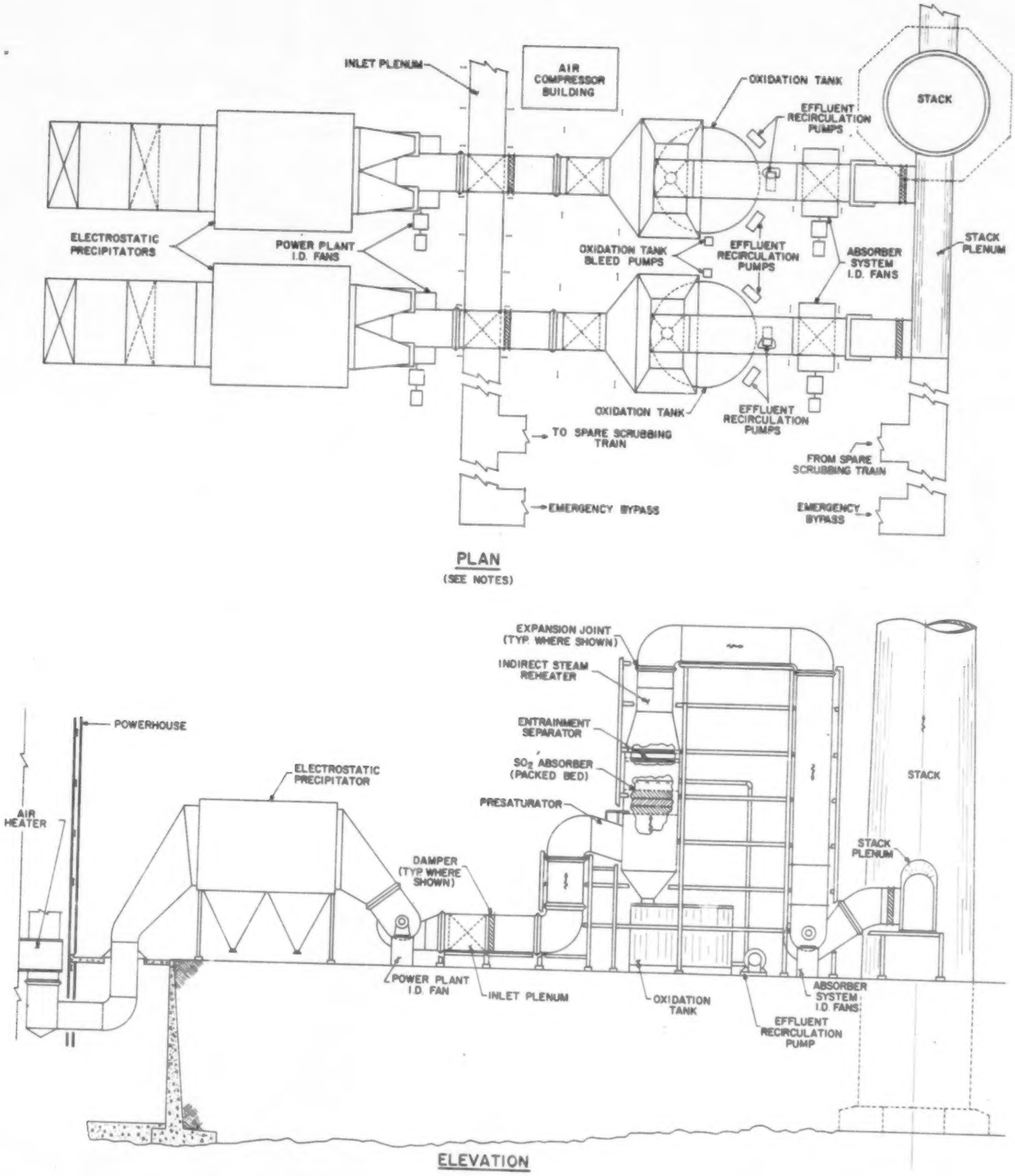
TABLE 17. (Continued)

	Stream No.	41				
	Description	CO_2 vent from 1st neutralization tank				
1	Total stream, lb/hr	9,686				
2						
3	Flow rate, sft^3/min(60°F)	1,393				
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gpm					
7	Specific gravity					
8	Basicity					
9	Undissolved solids, %					
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						


INLET PLENUM
AIR COMPRESSOR BUILDING
OXIDATION TANK
EFFLUENT RECIRCULATION PUMPS
STACK
ELECTROSTATIC PRECIPITATORS
POWER PLANT I.D. FANS
OXIDATION TANK BLEED PUMPS
EFFLUENT RECIRCULATION PUMPS
ABSORBER SYSTEM I.D. FANS
STACK PLENUM
OXIDATION TANK
EFFLUENT RECIRCULATION PUMPS
TO SPARE SCRUBBING TRAIN
FROM SPARE SCRUBBING TRAIN
EMERGENCY BYPASS
EMERGENCY BYPASS
PLAN
(SEE NOTES)
EXPANSION JOINT (TYP. WHERE SHOWN)
INDIRECT STEAM REHEATER
ENTRAINMENT SEPARATOR
SO₂ ABSORBER (PACKED BED)
PRESATURATOR
POWERHOUSE
ELECTROSTATIC PRECIPITATOR
AIR HEATER
STACK
DAMPER (TYP WHERE SHOWN)
STACK PLENUM
POWER PLANT I.D. FAN
INLET PLENUM
OXIDATION TANK
EFFLUENT RECIRCULATION PUMP
ABSORBER SYSTEM I.D. FANS
ELEVATION

NOTES:
A. EMERGENCY BYPASS ON EACH SIDE
B. SPARE SCRUBBING TRAIN ON ONE SIDE ONLY

Figure 10. Dowa process plan and elevation diagram.

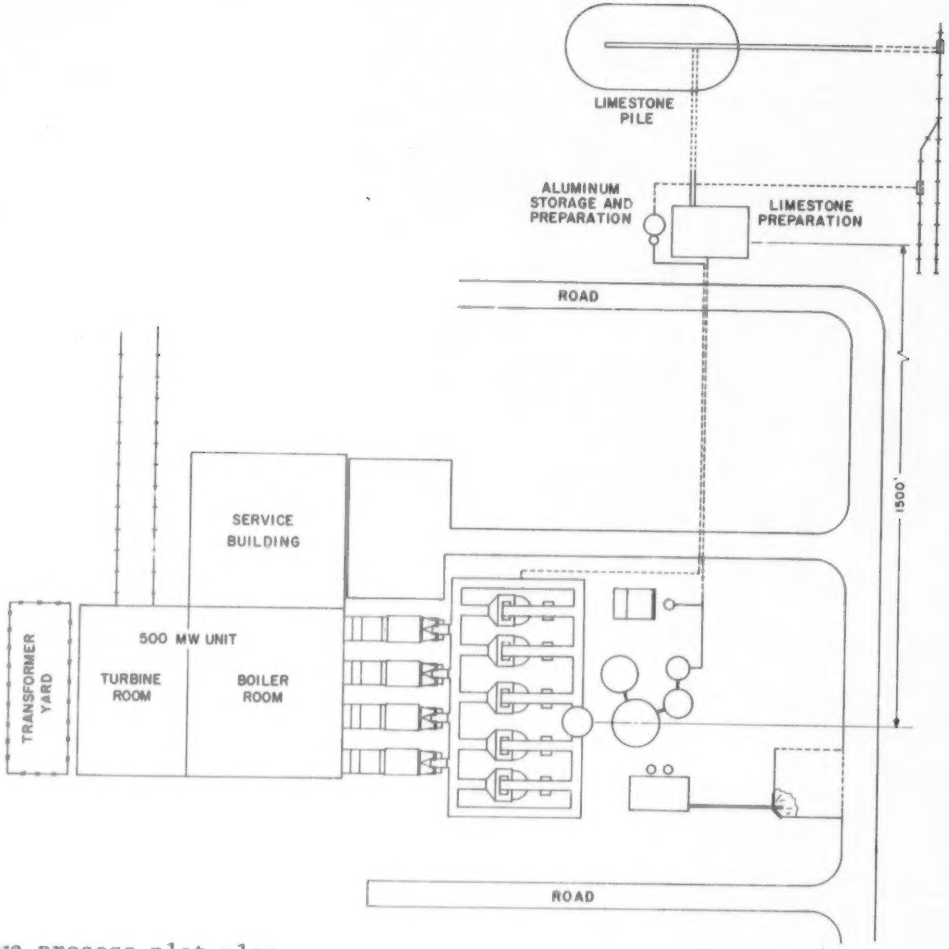

LIMESTONE PILE
ALUMINUM STORAGE AND PREPARATION
LIMESTONE PREPARATION
ROAD
1500'
SERVICE BUILDING
500 MW UNIT
TURBINE ROOM
BOILER ROOM
TRANSFORMER YARD
ROAD

Figure 11. Dowa process plot plan.

TABLE 18. DOWA PROCESS

EQUIPMENT LIST

Item (number): description	Total material cost, 1982 $
Area 1--Materials Handling	
1. Car shaker (1): Top mounting with crane, 20-hp shaker, 7-1/2-hp hoist	71,900
2. Car puller (1): 25-hp puller, 5-hp return	63,100
3. Hopper, limestone unloading (1): 16-ft diameter x 10-ft straight side, 2,570 ft3, carbon steel, 30-degree cone bottom	15,500
4. Feeder, limestone unloading (1): Vibrating pan, 3-1/2 hp, 250 tons/hr	5,500
5. Conveyor, limestone unloading (1): Belt, 36 in. wide x 20 ft long, 5 hp, 250 tons/hr, 165 ft/min	11,400
6. Conveyor, limestone unloading (incline) (1): Belt, 36 in. wide x 310 ft long, 50 hp, 15-degree incline, 80-ft total lift, 250 tons/hr, 165 ft/min	85,300
7. Dust collector, limestone unloading pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning, includes dust hood	11,200
8. Pump, limestone pit sump (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400
9. Conveyor, storage (1): Belt, 36 in. wide x 200 ft long, 5 hp, 250 tons/hr, 165 ft/min	73,100
10. Tripper, storage conveyor (1): 1 hp, 30 ft/min	27,200
11. Mobile equipment (1): Scraper tractor	141,900
12. Hopper, reclaim (2): 7-ft diameter x 4-1/4 ft deep x 2-ft bottom, 75 ft3, carbon steel, 60-degree cone bottom	2,400
13. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp, 100 tons/hr	10,900

(Continued)

TABLE 18. (Continued)

Item (number): description	Total material cost, 1982 $
14. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning	7,800
15. Pump, reclaim pit sump (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400
16. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 200 ft long, 5 hp, 100 tons/hr, 105 ft/min	40,900
17. Conveyor, limestone reclaim (incline) (1): Belt, 30 in. wide x 193 ft long, 40 hp, 15-degree incline, 50-ft total lift, 100 tons/hr, 105 ft/min	60,300
18. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90-ft lift, 100 tons/hr	57,800
19. Conveyor, live limestone feed (1): Belt, 30 in. wide x 60 ft long, 7-1/2 hp, 100 tons/hr, 105 ft/min	20,500
20. Tripper, live limestone feed conveyor (1): 1 hp, 30 ft/min	27,200
21. Bin, crusher feed (3): 13-ft diameter x 21-ft straight side height, 3,100 ft3, w/cover, 50-degree slope, carbon steel (2 operating, 1 spare)	43,300
22. Conveyor, aluminum sulfate unloading (1): Pneumatic, pressure system, 80 hp, 3,000 sft3/min, epoxy lined, includes bag filter on silo	64,800
23. Silo, aluminum sulfate storage (1): 22-ft diameter x 43-1/2-ft straight side height, 16,500 ft3, 60-degree slope, carbon steel, epoxy lined	60,000
24. Bin activator, aluminum sulfate storage silo (1): 12-ft diameter, 7-1/2 hp, carbon steel, epoxy lined	17,300
Subtotal	924,100
Area 2--Feed Preparation	
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp (2 operating, 1 spare)	49,600

(Continued)

TABLE 18. (Continued)

Item (number): description	Total material cost, 1982 $
2. Crusher (3): Gyratory, 75 hp (2 operating, 1 spare)	297,100
3. Ball mill (3): Wet, open system, 9-ft diameter x 16 ft long, 600 hp, 235 tons/day (2 operating, 1 spare)	1,632,300
4. Dust collector, ball mill (3): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning (2 operating, 1 spare)	23,300
5. Tank, mills product (3): 10-ft diameter x 10 ft high, 5,500 gal, open top, four 10-in. baffles, agitator supports, carbon steel, flakeglass lined (2 operating, 1 spare)	13,700
6. Agitator, mills product tank (3): 40-in. diameter, 10 hp, neoprene coated (2 operating, 1 spare)	22,900
7. Pump, mills product tank (3): Centrifugal, 40 gpm, 60-ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,400
8. Tank, slurry feed (1): 19-ft diameter x 19 ft high, 40,450 gal, open top, four 19-in. baffles, agitator supports, carbon steel, flakeglass lined	16,500
9. Agitator, slurry feed tank (1): 76-in. diameter, 40 hp, neoprene coated	33,600
10. Pump, slurry feed tank (8): Centrifugal, 20 gpm, 60-ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spares)	21,400
11. Feeder, aluminum sulfate (1): Weigh belt, 20 ft long, 316 stainless steel, 2 hp, with dust enclosure	29,800
12. Tank, aluminum sulfate mixing (1): 9-1/2-ft diameter x 9 ft high, 4,800 gal, open top, four 9-in. baffles, agitator supports	3,700
13. Agitator, aluminum sulfate mixing tank (1): 38-in. diameter, 5 hp, neoprene coated	3,400

(Continued)

TABLE 18. (Continued)

Item (number): description	Total material cost, 1982 $
14. Pump, aluminum sulfate mixing tank (6): Centrifugal, 2-1/2 gpm, 60-ft head, 1/2 hp, 316 stainless steel (4 operating, 2 spares)	9,000
Subtotal	2,164,700
Area 3--Gas Handling	
1. Fans (5): Induced draft, 379,100 aft^3/min, 7.6-in. static head, 700 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	3,380,200
Subtotal	3,380,200
Area 4--SO$_2$ Absorption	
1. SO$_2$ absorber (5): Packed bed, 40 ft high x 39 ft wide x 20 ft deep, 1/4-in. 316 stainless steel, 316 stainless steel grids, FRP chevron vane entrainment separator, fixed-bed polygrid packing (4 operating, 1 spare)	6,463,000
2. Pump, effluent recirculation (15): Centrifugal, 7,350 gpm, 100-ft head, 400 hp, 316 stainless steel (8 operating, 7 spares)	1,402,000
3. Pump, makeup water (2): Centrifugal, 625 gpm, 200-ft head, 60 hp, carbon steel (1 operating, 1 spare)	19,600
4. Sootblowers (40): Air, fixed	111,900
Subtotal	7,996,500
Area 5--Reheat	
1. Reheater (5): Steam, tube type, 3,000 ft^2, one-half of tubes made of Inconel 625 and one-half made of Cor-Ten (4 operating, 1 spare)	2,727,900
2. Sootblowers (20): Air, retractable	183,000
Subtotal	2,910,900

(Continued)

TABLE 18. (Continued)

Item (number): description	Total material cost, 1982 $
Area 6--Oxidation	
1. Tank, oxidation (5): 33-1/2-ft diameter x 17 ft high, 112,000 gal, open top, four 33-1/2-in.-wide baffles, agitator supports, 316 stainless steel, includes 316 stainless steel sparger (4 operating, 1 spare)	493,600
2. Agitator, oxidation tank (5): 134-in. diameter, 75 hp, 316 stainless steel shaft and turbine (4 operating, 1 spare)	481,100
3. Pump, oxidation tank bleed (8): Centrifugal, 972 gpm, 60-ft head, 30 hp, 316 stainless steel (4 operating, 4 spares)	112,700
4. Blower, oxidation air (6): 5,279 sft3/min, 500 hp (4 operating, 2 spares)	337,200
Subtotal	1,424,600
Area 7--Neutralization	
1. Tank, redissolution (1): 28-1/2-ft diameter x 28-1/2 ft high, 135,600 gal, open top, four 28-1/2-in.-wide baffles, agitator supports, carbon steel, flakeglass lined	35,200
2. Agitator, redissolution tank (1): 114-in. diameter, 40 hp, neoprene coated	49,300
3. Tank, 1st neutralization (1): 33-ft diameter x 28-1/2 ft high, 182,400 gal, open top, four 33-in.-wide baffles, agitator supports, carbon steel, flakeglass lined	41,900
4. Agitator, 1st neutralization tank (1): 132-in. diameter, 50 hp, neoprene coated	62,300
5. Pump drain (1): Centrifugal, 1,250 gpm, 50-ft head, 40 hp, 316 stainless steel (1 spare)	15,600
Subtotal	204,300

(Continued)

TABLE 18. (Continued)

Item (number): description	Total material cost, 1982 $
Area 8--Chloride Purge	
1. Tank, 2nd neutralization (1): 11-1/2-ft diameter x 11-1/2 ft high, 8,900 gal, open top, four 11-1/2-in.-wide baffles, agitator supports, carbon steel, flakeglass lined	5,700
2. Agitator, 2nd neutralization tank (1): 46-in. diameter, 7-1/2 hp	7,500
3. Pump, 2nd neutralization tank (2): Centrifugal, 287 gpm, 50-ft head, 7-1/2 hp, carbon steel, neoprene lined (1 operating, 1 spare)	8,700
4. Tank, settling (1): 35 ft wide x 49 ft long x 25 ft high, 221,800 gal, open top, 30-degree sloped bottom in first 36 ft of tank, 2 internal baffles, agitator supports, carbon steel, flakeglass lined	60,100
5. Agitator, settling tank (3): 52-in. diameter, 30 hp	50,100
6. Pump, chloride purge (2): Centrifugal, 65 gpm, 125-ft head, 3-1/2 hp, carbon steel, neoprene lined (1 operating, 1 spare)	6,100
7. Pump, effluent return (2): Centrifugal, 230 gpm, 75-ft head, 10 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,200
Subtotal	147,400
Area 9--Solids Separation	
1. Thickener (1): 55-ft diameter x 5-1/2 ft high, carbon steel sides, concrete basin, includes 1-1/2-hp rake and motor mechanism	89,400
2. Tank, thickener overflow (1): 46-ft diameter x 5-1/2 ft high, 68,000 gal, open top, carbon steel, flakeglass lined	33,000
3. Pump, thickener overflow tank (2): Centrifugal, 4,121 gpm, 75-ft head, 200 hp, carbon steel, neoprene lined (1 operating, 1 spare)	61,200

(Continued)

TABLE 18. (Continued)

Item (number): description	Total material cost, 1982 $
4. Pump, thickener underflow (2): Centrifugal, 1,621 gpm, 10-ft head, 10 hp, carbon steel, neoprene lined (1 operating, 1 spare)	14,500
5. Tank, filter feed (1): 9-ft diameter x 8-1/2 ft high, 4,050 gal, open top, four 9-in.-wide baffles, agitator supports, carbon steel, flakeglass lined	3,400
6. Agitator, filter feed tank (1): 36-in. diameter, 7-1/2 hp, neoprene coated	5,000
7. Pump, filter feed tank (3): Centrifugal, 122 gpm, 50-ft head, 5 hp, carbon steel, neoprene lined (2 operating, 1 spare)	12,100
8. Filter (3): Rotary vacuum, 10-ft diameter x 10-ft face, 50 total hp, includes auxiliary equipment (2 operating, 1 spare)	345,600
9. Pump, filtrate (4): Centrifugal, 118 gpm, 20-ft head, 1-1/2 hp, carbon steel, neoprene lined (2 operating, 2 spares)	14,900
10. Tank, filtrate surge (1): 9-ft diameter x 8-1/2 ft high, 4,050 gal, open top, carbon steel, flakeglass lined	3,100
11. Pump, filtrate surge tank (2): Centrifugal, 236 gpm, 85-ft head, 10 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,200
12. Conveyor, filter cake (1): Belt, 30 in. wide x 180 ft long, 1-1/2 hp, 40 tons/hr	38,000
13. Retaining wall, gypsum (1): 8 ft high x 32 ft long, concrete	1,600
Subtotal	631,000

third spare ball mill is provided. The 60% solids slurry is stored in the slurry feed tank (8-hour residence time), situated near the absorber system about 1,500 feet from the limestone preparation area.

Aluminum sulfate is metered by a weigh belt feeder into the solution tank. The solution is then pumped into the oxidation tank for aluminum makeup.

Gas Handling--

The inlet plenum distributes the flue gas to the five ducts (base case) which feed the absorbers. An emergency bypass duct designed to handle up to 50% of the boiler flue gas is included as two 25% capacity ducts, one around each end of the facility. An ID fan after the reheater compensates for the 7.6-inch H_2O pressure drop (base case) through the FGD system. The clean gas is then ducted to the stack plenum and stack. All flue gas ductwork between the power unit ID fans and the stack plenum is included in this area.

SO_2 Absorption--

Upon entering the absorber, the flue gas passes through a fresh water wall wash which performs two tasks: it prevents solids buildup at the gas-liquid interface by constantly washing the area clean, and it cools the entering gas from 300°F to 125°F to prevent damage to the packing and other absorber internals. The absorber is packed with Ecodyne polygrid packing, a plastic modular-type packing with an open area of about 80%. It is supplied in 2-foot x 4-foot x 1-1/4-inch sections which are stacked on supported grids to a depth of 8 feet. The basic aluminum sulfate solution, at a basicity of 10%, is pumped from the oxidation tank to the packed tower and is contacted countercurrently with the gas to remove SO_2. The system is operated at a total stoichiometry (including neutralization and chloride purge) of 1.03 mols of $CaCO_3$/mol (SO_2 + 2HCl) absorbed with a L/G ratio of 45 gal/kaft^3. Chevron vane entrainment separators are used to reduce liquid carryover.

Reheat--

An indirect steam reheater, equipped with sootblowers for cleaning, is provided in each train. The temperature of the cleaned gas is increased to 170°F to evaporate entrained moisture and increase plume buoyancy before the gas enters the ID fan and stack. The heat of compression in the ID fan increases the flue gas temperature to 175°F at the stack.

Oxidation--

One oxidation tank per train collects the scrubber effluent containing an aluminum sulfite/sulfate complex. The oxidation tank has a liquid level of 16 feet above the sparger. A blower serving each oxidation tank forces air into each tank through spargers at an oxidation efficiency of 25% (4.0 atoms O/mol SO_2). [Current research work by UOP expects to support Japanese Dowa experience of 50% oxidation efficiency (103).] Makeup aluminum sulfate from the solution tank is added to the system through the oxidation tank.

Neutralization--

A bleedstream from the oxidation tank is pumped to the redissolution tank (30-minute residence time) along with a stream from the settling tank (see the Chloride Purge area). The redissolution tank overflows into the first neutralizing tank (30-minute residence time) where limestone is added to regenerate the aluminum sulfate solution to a basicity of 25% and precipitate gypsum. (UOP suggests a regenerated solution of 30% basicity.)

Chloride Purge--

A purge from the thickener overflow stream is pumped into the second neutralization tank (30-minute residence time) where limestone is added to precipitate gypsum and aluminum hydroxide (for a more detailed explanation see the Dowa process description section). The slurry is pumped into a specially designed settling tank to separate the solids from the aluminum-free solution. Chlorides are removed by purging a small amount of the solution to the fly ash pond. The underflow, containing precipitated aluminum hydroxide and gypsum, is pumped to the redissolution tank where the aluminum hydroxide is dissolved for reuse.

Solids Separation--

Overflow from the first neutralization tank (12% solids) flows into a thickener where the gypsum crystals are concentrated to 35% solids and further dewatered to 85% solids in a rotary vacuum filter. (UOP standard design utilizes a 10% solids overflow from the first neutralization tank and an 85% solids filter cake.) The filtrate is combined with the thickener overflow and returned to the oxidation tank. The dewatered gypsum is conveyed to an in-process storage area where wheeled loaders transfer the gypsum to dump trucks for transport to a landfill situated one mile from the scrubbing facilities. The landfill is described in the premises. For the base case, the landfill is square with sides of approximately 1,770 feet and covering an area of 104 acres.

RESULTS

Based on the power plant design and economic premises and the specific process designs as described in the previous sections, capital investments, first-year annual revenue requirements, and levelized annual revenue requirements were prepared for the economic evaluation and comparison of the advanced limestone, Davy S-H, and Dowa FGD processes.

In addition to the base case 500-MW new power unit (30-year lifetime) burning a 3.5% sulfur bituminous coal, case variations for different power unit sizes (200 and 1,000 MW), remaining life (20 years), and coal sulfur contents (2.0% and 5.0%) are determined. Sensitivity analyses were also performed to evaluate the effects of varying some of the major cost elements in the annual revenue requirements that are subject to geographical and economic influences.

Detailed capital investment and annual revenue requirement tables for each base case and case variation are shown in the appendix. Capital investment is based on mid-1982 costs. Annual revenue requirements are based on a one-year, 5,500-hour full-load operation and mid-1984 costs.

CAPITAL INVESTMENT

The capital investments for each process are summarized in Table 19. In the base case comparison, the advanced limestone process has the lowest capital investment, followed by the Davy S-H process, 25% higher, and the Dowa process, 31% higher.

Base Case

A breakdown of the total direct capital investment by area for the base cases is shown in Table 20. The materials-handling area cost for the Dowa process is slightly higher than it is for the advanced limestone process because of the added cost of the aluminum sulfate addition system. Other than this, the costs in this area are equivalent for these two processes. The Davy S-H process, which uses lime and additives, has the highest materials-handling cost. Although the materials-handling equipment cost (from the equipment list) is lowest for the Davy S-H process, the greater cost of ancillary equipment such as piping, electrical equipment, instrumentation, etc., more than makes up the difference, giving the Davy S-H process the highest materials-handling cost.

Feed preparation area cost for the Davy S-H process is only one-fourth those of the other two processes because of the simpler feed system used for a lime-based system, which does not require crushing and milling equipment. The advanced limestone and Dowa processes are essentially equivalent in feed preparation area cost except for the Dowa aluminum sulfate addition cost.

TABLE 19. CAPITAL INVESTMENT SUMMARY

	Advanced limestone		Davy S-H		Dowa	
	M$	$/kW	M$	$/kW	M$	$/kW
New Units[a]						
200 MW, 3.5% S	54.4	272	63.0	315	66.4	332
500 MW, 2.0% S	79.1	158	99.8	200	102.6	205
500 MW, 3.5% S	92.8	186	115.5	231	121.3	243
500 MW, 5.0% S	97.9	196	121.4	243	130.3	261
1,000 MW, 3.5% S	155.5	156	206.7	207	216.1	216
Existing Units[b]						
200 MW, 3.5% S	59.8	299	80.6	403	85.4	427
500 MW, 3.5% S	118.5	237	148.1	296	156.2	312
1,000 MW, 3.5% S	211.7	212	268.1	268	281.3	281

a. New FGD facilities are constructed simultaneously with the power plant and have a 30-year remaining life.

b. Existing FGD facilities are retrofit installations on existing power plant facilities and have a 20-year remaining life.

Gas-handling cost, being related to the system pressure drop, is highest for the Davy S-H process because of the high pressure drop in the venturi absorber and the high duct cost. The high cost for ducts in the Davy S-H process is a result of the large amount of 316 stainless steel ducting required between the Rotoparts and the reheaters, where the wet acidic gas is below 150°F. The Dowa process, which uses a packed-bed absorber, has the next highest pressure drop, and therefore, the next highest area cost.

The SO_2 absorption areas are not readily comparable because of process design differences, but some relevant points can be mentioned. The advanced limestone process has a slightly higher cost since it has a higher L/G ratio, which increases the size of liquid-handling equipment compared with the Dowa and Davy S-H processes. The Davy S-H process has the lowest SO_2 absorption area cost because of the simplicity of its modular absorbers. The reheat area investment cost is substantially lower for the advanced limestone process because of the reheat benefit received from partial scrubbing. Reheat area costs are essentially equivalent for the Davy S-H and Dowa processes since all cases have the same number of absorption trains and similar flue gas quantities and inlet and outlet temperatures.

TABLE 20. CAPITAL INVESTMENT BY PROCESSING AREAS FOR NEW

500-MW POWER UNITS WITH 3.5% SULFUR COAL

	M$		
	Advanced limestone	Davy S-H	Dowa
Materials handling	2.6	3.5	2.9
Feed preparation	4.3	1.2	5.3
Gas handling	10.5	18.6	14.3
SO_2 absorption	17.7	14.7	17.6
Reheat	3.2	4.2	4.0
Oxidation	2.3	5.0	4.7
Neutralization	-	-	0.6
Chloride purge	-	-	0.8
Solids separation	3.5	2.6	2.7
Total process capital	44.1	49.8	52.9
Services, utilities, and miscellaneous	2.6	3.0	3.2
Landfill construction	3.4	3.3	3.3
Landfill equipment	0.9	0.9	0.9
Total direct investment	51.0	57.0	60.3
Total capital investment	92.8	115.5	121.3

The oxidation area costs are highest for the Davy S-H process because of a unique air sparger design. The Dowa process, slightly lower in cost, has a stainless steel sparger of standard design. Stainless steel must be used because of the corrosive nature of the effluent. The advanced limestone process has the *lowest oxidation area cost* partially because of the lower air stoichiometry which reduces blower costs.

The neutralization area and chloride purge area costs of the Dowa process contribute very little (less than 2% each) to the total direct capital investment.

The solids separation area cost is highest for the advanced limestone process. The costs for ancillary equipment, primarily electrical equipment and piping, are greater for the advanced limestone process because of the higher flow rates in this area. This area does not include costs associated with the landfill such as construction costs and equipment. All landfill costs are listed separately and are a function of gypsum production rate and lifetime gypsum volume. Since all processes produce approximately the same quantity of gypsum, the capital investments for landfill are similar.

The areas that have the most effect on the order of the base case capital investments are the feed preparation, gas-handling, SO_2 absorption, reheat, and oxidation areas. The high gas-handling and oxidation area costs of the Davy S-H process, compared with the advanced limestone process, are not offset by its lower feed preparation and SO_2 absorption area costs and result in the Davy S-H process having the second highest capital investment. The Dowa process, with only two area costs (SO_2 absorption and solids separation) lower than the advanced limestone process, has the highest base case capital investment.

Case Variations

The capital investments of 200-, 500-, and 1,000-MW power units are shown in Table 21 for new units and in Table 22 for existing units. The same comparisons are shown graphically in Figures 12 and 13. The process ranking, in order of capital investment cost, remains the same for the 200- and 1,000-MW units as it was for the 500-MW base case. Capital investments approximately double as the power unit size increases from 200 to 500 MW and triple as the size increases from 200 to 1,000 MW. There is a greater economy of scale than may be expected from the 200- to 500-MW power unit size as a result of the percentage of spare absorber capacity. The 200-MW units are provided with 50% spare capacity--one spare with two operating trains--while the 500- and 1,000-MW units have 25% spare capacity--one spare for each four operating trains.

The capital investments for new 500-MW units burning 2.0%, 3.5%, and 5.0% sulfur coal are shown in Table 23 and Figures 14 and 15. Compared to the 3.5% sulfur base case, the capital investments for the 2.0% sulfur coal are about 15% lower for all three processes and for the 5.0% sulfur coal they are 5% to 7% higher.

ANNUAL REVENUE REQUIREMENTS

Detailed annual revenue requirements for each process are shown in the appendix and are summarized in Table 24. The annual revenue requirement relationship between processes differs from that of the capital investment. In the base case comparison, the Davy S-H process has the highest annual revenue requirements, followed in decreasing order by the Dowa process and the advanced limestone process.

TABLE 21. CAPITAL INVESTMENT BY PROCESSING AREA FOR NEW 200-, 500-, AND 1,000-MW POWER UNITS WITH 3.5% SULFUR COAL

	M$								
	Advanced limestone			Davy S-H			Dowa		
	200	500	1,000	200	500	1,000	200	500	1,000
Materials handling	1.4	2.6	2.9	1.9	3.5	5.4	1.3	2.9	3.9
Feed preparation	2.8	4.3	5.6	0.7	1.2	1.9	3.0	5.3	9.2
Gas handling	5.5	10.5	19.8	9.7	18.6	36.4	7.5	14.3	28.0
SO_2 absorption	9.5	17.7	34.0	7.3	14.7	28.5	9.0	17.6	34.3
Reheat	1.7	3.2	6.2	2.2	4.2	8.1	2.1	4.0	7.7
Oxidation	1.4	2.3	4.4	2.6	5.0	9.9	2.5	4.7	9.3
Neutralization	-	-	-	-	-	-	0.3	0.6	1.1
Chloride purge	-	-	-	-	-	-	0.5	0.8	1.2
Solids separation	2.7	3.5	4.4	1.7	2.6	3.5	1.6	2.7	4.0
Total process capital	25.0	44.1	77.3	26.1	49.8	93.7	27.8	52.9	98.7
Services, utilities, and miscellaneous	1.5	2.6	4.7	1.6	3.0	5.6	1.7	3.2	6.0
Landfill construction	1.9	3.4	5.2	1.8	3.3	5.1	1.8	3.3	5.1
Landfill equipment	0.5	0.9	1.5	0.5	0.9	1.5	0.5	0.9	1.5
Total direct investment	28.9	51.0	88.7	30.0	57.0	105.9	31.8	60.3	111.3
Total capital investment	54.4	92.8	155.5	63.0	115.5	206.7	66.4	121.3	216.1

TABLE 22. CAPITAL INVESTMENT BY PROCESSING AREA FOR EXISTING 200-, 500-, AND 1,000-MW POWER UNITS WITH 3.5% SULFUR COAL

	M$								
	Advanced limestone			Davy S-H			Dowa		
	200	500	1,000	200	500	1,000	200	500	1,000
Materials handling	1.9	3.4	5.2	2.5	4.6	7.2	1.8	3.3	5.2
Feed preparation	3.0	5.7	9.2	0.9	1.6	2.5	4.0	7.6	12.3
Gas handling	6.4	13.8	27.2	12.8	24.6	48.4	9.9	18.8	37.2
SO_2 absorption	10.6	23.4	46.1	9.7	19.4	38.1	11.9	23.2	45.6
Reheat	1.9	4.2	8.4	2.8	5.5	10.8	2.7	5.2	10.3
Oxidation	1.4	3.1	6.1	3.5	6.6	13.1	3.3	6.3	12.3
Neutralization	-	-	-	-	-	-	0.4	0.7	1.4
Chloride purge	-	-	-	-	-	-	0.6	1.1	1.6
Solids separation	2.8	4.5	6.6	2.2	3.4	4.7	2.1	3.6	5.4
Total process capital	28.0	58.1	108.8	34.4	65.7	124.8	36.7	69.8	131.3
Services, utilities, and miscellaneous	1.7	3.5	6.5	2.1	4.0	7.5	2.2	4.2	7.9
Landfill construction	1.5	2.6	4.1	1.4	2.5	3.9	1.4	2.5	4.0
Landfill equipment	0.5	0.9	1.5	0.5	0.9	1.5	0.5	0.9	1.5
Total direct investment	31.7	65.1	120.9	38.4	73.0	137.7	40.8	77.4	144.7
Total capital investment	59.8	118.5	211.7	80.6	148.1	268.1	85.4	156.2	281.3

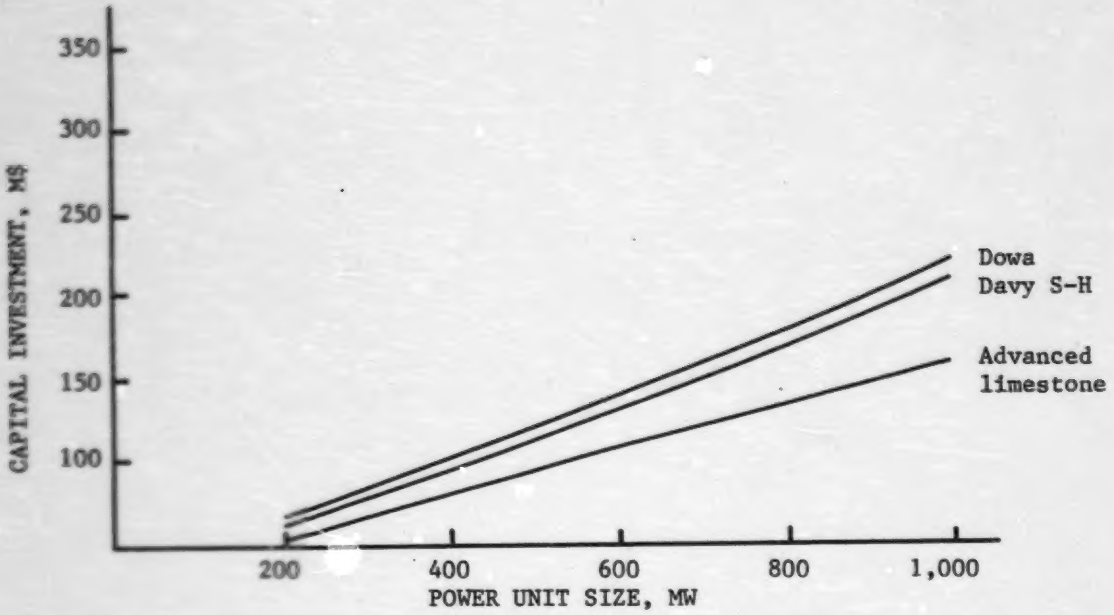


Figure 12. Power unit size versus total capital investment for new units.

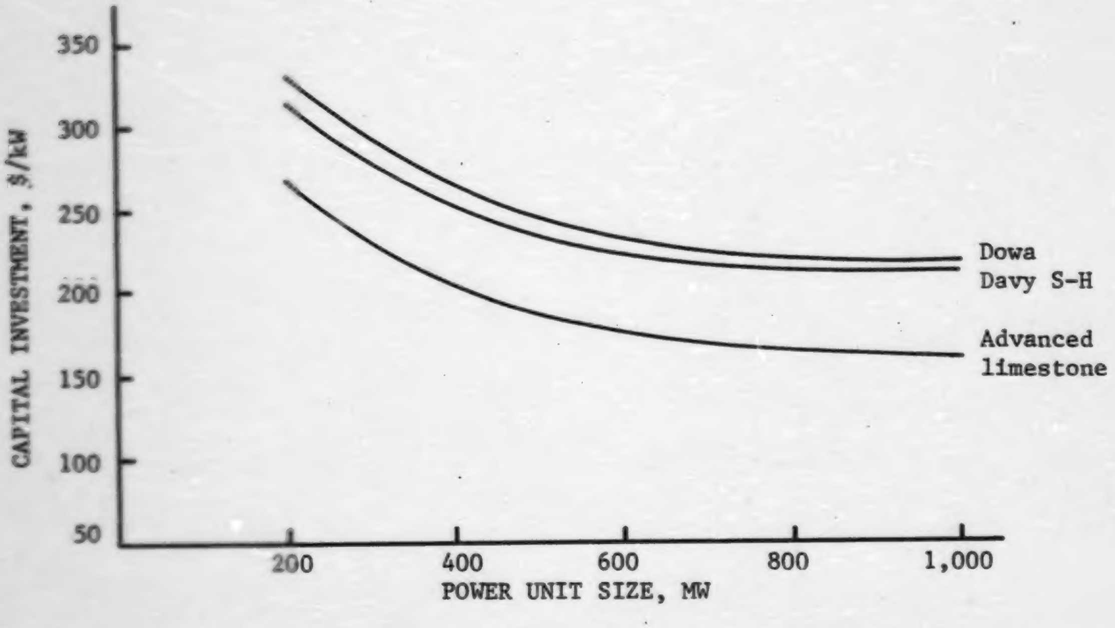


Figure 13. Power unit size versus unit capital investment for new units.

TABLE 23. CAPITAL INVESTMENT BY PROCESSING AREA FOR NEW

500-MW POWER UNITS WITH 2.0%, 3.5%, AND 5.0% SULFUR COAL

	M$								
	Advanced limestone			Davy S-H			Dowa		
	2.0	3.5	5.0	2.0	3.5	5.0	2.0	3.5	5.0
Materials handling	1.3	2.6	2.7	2.2	3.5	4.4	1.6	2.9	3.2
Feed preparation	2.8	4.3	4.9	0.8	1.2	1.6	3.6	5.3	7.4
Gas handling	9.8	10.5	10.5	17.2	18.6	18.6	13.2	14.3	14.3
SO_2 absorption	16.7	17.7	17.8	13.2	14.7	14.6	16.1	17.6	17.4
Reheat	2.6	3.2	3.2	3.8	4.1	4.2	3.6	4.0	4.0
Oxidation	2.3	2.3	2.6	4.7	5.0	5.0	4.4	4.7	4.7
Neutralization	-	-	-	-	-	-	0.5	0.6	0.6
Chloride purge	-	-	-	-	-	-	0.6	0.8	1.0
Solids separation	2.9	3.5	3.9	1.9	2.7	3.0	1.8	2.7	3.4
Total process capital	38.4	44.1	45.6	43.8	49.8	51.4	45.4	52.9	56.0
Services, utilities, and miscellaneous	2.3	2.6	2.7	2.6	3.0	3.1	2.7	3.2	3.4
Landfill construction	2.2	3.4	4.3	2.1	3.3	4.2	2.1	3.3	4.2
Landfill equipment	0.6	0.9	1.3	0.6	0.9	1.2	0.6	0.9	1.3
Total direct investment	43.5	51.0	53.9	49.1	57.0	59.9	50.8	60.3	64.9
Total capital investment	79.1	92.8	97.9	99.8	115.5	121.4	102.6	121.3	130.3

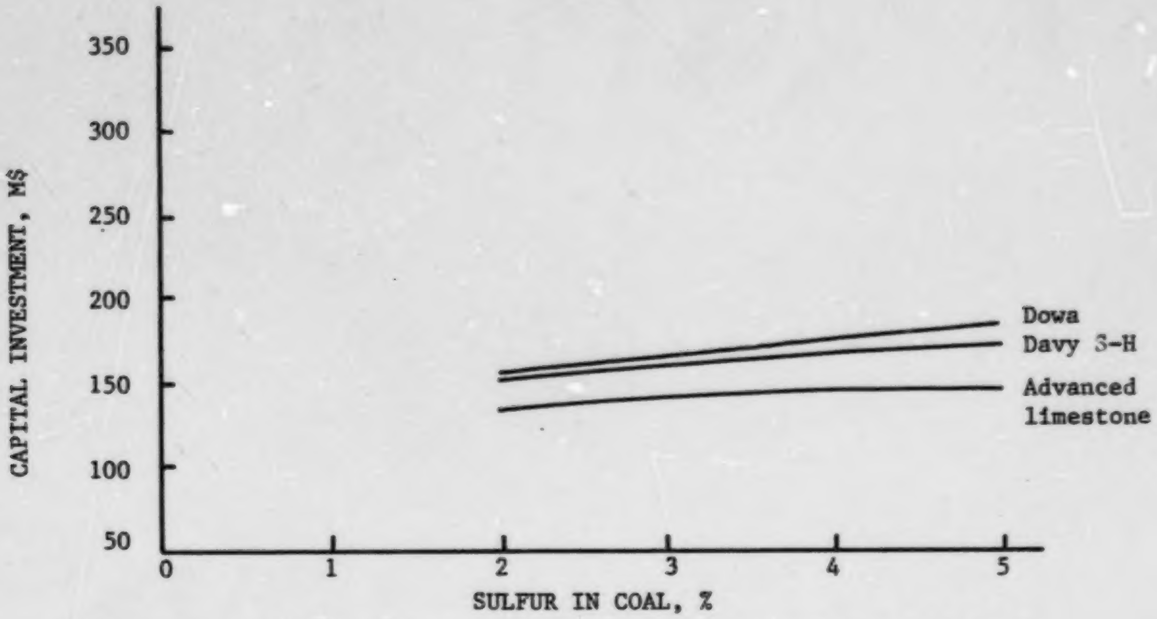


Figure 14. Sulfur in coal versus total capital investment.

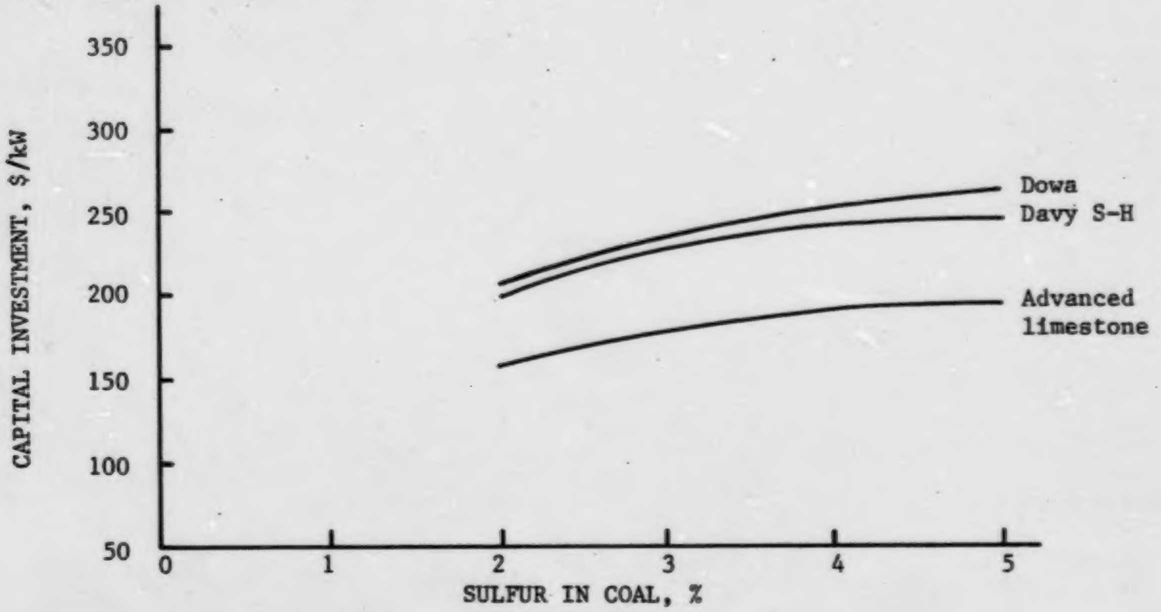


Figure 15. Sulfur in coal versus unit capital investment.

TABLE 24. ANNUAL REVENUE REQUIREMENT SUMMARY

	Advanced limestone				Davy S-H				Dowa			
	First year		Levelized		First year		Levelized		First year		Levelized	
	M$	Mills/kWh	M$	Mills/kWh	M$	Mills/kWh	M$	Mills/kWh	M$	Mills/kWh	M$	Mills/kWh
New Units[a]												
200 MW, 3.5% S	15.3	13.95	21.9	19.86	17.7	16.06	25.1	22.83	18.0	16.33	25.2	22.94
500 MW, 2.0% S	21.6	7.86	30.4	11.07	26.7	9.69	37.3	13.55	26.8	9.76	37.2	13.54
500 MW, 3.5% S	25.8	9.38	36.6	13.31	32.7	11.88	46.5	16.93	32.3	11.74	45.1	16.39
500 MW, 5.0% S	27.7	10.07	39.5	14.36	35.9	13.08	52.0	18.90	34.8	12.65	48.6	17.68
1,000 MW, 3.5% S	42.8	7.78	60.4	10.99	57.4	10.44	81.4	14.80	56.1	10.21	77.7	14.13
Existing Units[b]												
200 MW, 3.5% S	17.1	15.53	22.0	20.04	21.9	19.87	27.8	25.28	22.7	20.66	28.7	26.14
500 MW, 3.5% S	32.5	11.83	41.5	15.10	40.2	14.63	51.2	18.62	40.9	14.86	51.4	18.71
1,000 MW, 3.5% S	56.9	10.34	72.1	13.12	71.6	13.02	90.6	16.48	71.9	13.08	90.0	16.36

a. New FGD facilities are constructed simultaneously with the power plant and have a 30-year remaining life.

b. Existing FGD facilities are retrofit installations on existing power plant facilities and have a 20-year remaining life.

Base Case

The differences in annual revenue requirements among these processes lie in the differences in the costs of raw materials, electricity, maintenance, capital charges, and overheads. These components are shown in order of magnitude for each process in Table 25. Other costs are either insignificant or similar for all three processes and therefore, have no differential effect on the process comparison. With the exception of raw material cost, all processes have the same order of major cost components: levelized capital charges are the largest component, followed in order by maintenance, plant and administrative overheads, and electricity. Raw material cost is the second highest component of annual revenue requirements for the Davy S-H process because lime is used, whereas, it is the fifth highest component for the advanced limestone and Dowa processes. Capital charges are the largest component of the annual revenue requirements, consisting of over 50% of the total for all processes.

In terms of percentages, the contribution of capital charges to annual revenue requirements depends on the magnitude of both the capital investment and the operating and maintenance costs. The Davy S-H process capital charges are about 25% higher than the advanced limestone process capital charges. The Davy S-H process operating and maintenance costs are higher than those of the advanced limestone process by the same percentage, however, and capital charges constitute about 52% of the annual revenue requirements of both processes. In contrast, the Dowa process capital charges are only 5% higher than those of the Davy S-H process, but the Dowa process operating and maintenance costs are 9% lower. The Dowa process capital charges thus constitute about 55% of the annual revenue requirements. The significance of the percentage of annual revenue requirements that consists of capital charges lies in comparisons of levelized annual revenue requirements.

The levelized annual revenue requirements of processes with a higher percentage of capital charges in their annual revenue requirements (comprising either processes that are expensive to install or inexpensive to operate) differ less from first-year annual revenue requirements than those of processes in which capital charges constitute a lower percentage of the annual revenue requirements. The Dowa process first-year annual revenue requirements are only 1% lower than those of the Davy S-H process, but the levelized annual revenue requirements are 3% lower. Since levelized costs are based on long-term projections of economic conditions, the validity of comparisons using levelized costs are contingent on the accuracy of these projections. In this study, the advanced limestone process (with the lowest capital investment and lowest operating and maintenance costs) would be least affected by changes in the levelizing factor. Its cost relative to the Davy S-H and Dowa processes would be improved by increasing inflation and discount rates while the reverse

TABLE 25. MAJOR COST COMPONENTS FOR BASE CASE ANNUAL REVENUE REQUIREMENTS

	Major operating cost components (% of first-year annual revenue requirements)				
Process	1	2	3	4	5
Advanced limestone	Levelized capital charges 52.7	Maintenance 14.7	Plant and adm. overheads 12.0	Electricity 6.6	Raw materials 4.3
Davy S-H	Levelized capital charges 52.0	Raw materials 14.1	Maintenance 10.1	Plant and adm. overheads 8.3	Electricity 6.7
Dowa	Levelized capital charges 55.3	Maintenance 14.3	Plant and adm. overheads 10.9	Electricity 6.2	Raw materials 4.3

would occur with decreasing inflation and discount rates, in terms of levelized costs. The same effect would occur, to a lesser extent, for the Dowa process relative to the Davy S-H process.

The Davy S-H process has the highest raw material cost because of the use of lime at 75 $/ton, as compared with 8.50 $/ton for the limestone used in the advanced limestone and Dowa processes. The Davy S-H process raw material cost is 4.0 and 3.3 times those of the advanced limestone and Dowa processes, respectively. This significantly higher raw material cost is the reason that the Davy S-H process (with the lowest conversion cost) has the highest operating and maintenance costs and the highest annual revenue requirements.

Electricity cost depends on the quantity and size of equipment and process complexity. Electricity use is greatest in all cases for the Davy S-H process. However, in the most extreme case, the electricity use is only one-third greater than that of the other processes. Maintenance cost is based on direct process investment and depends on the size and complexity of process equipment. The plant and administrative overheads are closely related to maintenance and consequently, also to the direct process investment. For all cases, the Dowa process is assessed the highest charge for maintenance and plant and administrative overheads because of its higher direct process investment.

It is evident that capital investment cost plays a large role in the annual revenue requirements. Capital charges are a direct function of capital investment, maintenance is a function of direct process investment, and plant and administrative overheads include 60% of the maintenance charge. Therefore, the ranking of annual revenue requirements is somewhat inherited from the total capital investment ranking unless a cost component is large enough to overcome the effect, such as the Davy S-H raw material cost.

Case Variations

Annual revenue requirements for 200-, 500-, and 1,000-MW power units are shown in Table 26 for new units and in Table 27 for existing units. The same comparisons for new units are shown graphically in Figure 16. The levelized annual revenue requirements for new units increase 70% to 85% as the power unit size increases from 200 to 500 MW and 175% to 225% as the power unit size increases from 200 to 1,000 MW. Existing unit costs increase 80% to 90% for the 200- to 500-MW power unit size increase and 215% to 230% for the 200- to 1,000-MW increase.

The Davy S-H process has the higher levelized annual revenue requirements at the 500- and 1,000-MW power unit sizes, but at the 200-MW size, the levelized annual revenue requirements for the Dowa process are slightly higher than those of the Davy S-H process for new power units. This trend is related to changes in the percentages of the major cost components as the power unit size increases. Since the Dowa process has the highest direct process investment and total capital investment in every case, its maintenance

TABLE 26. ANNUAL REVENUE REQUIREMENT SUMMARY FOR NEW 200-, 500-, AND 1,000-MW POWER UNITS WITH 3.5% SULFUR COAL

	MS								
	Advanced limestone			Davy S-H			Dowa		
	200	500	1,000	200	500	1,000	200	500	1,000
Raw materials									
Limestone	0.4	0.9	1.8	-	-	-	0.4	0.9	1.8
Lime	-	-	-	1.8	4.3	8.4	-	-	-
Adipic acid	0.1	0.2	0.4	-	-	-	-	-	-
Formic acid	-	-	-	0.1	0.2	0.3	-	-	-
Ferric chloride	-	-	-	<0.1	<0.1	0.1	-	-	-
Antifoulant	-	-	-	<0.1	<0.1	<0.1	-	-	-
Aluminum sulfate	-	-	-	-	-	-	0.2	0.5	0.9
Total raw materials	0.5	1.1	2.2	1.9	4.6	8.8	0.6	1.4	2.7
Total labor	1.0	1.2	1.6	1.0	1.2	1.5	1.0	1.2	1.5
Utilities									
Electricity	0.7	1.7	3.2	0.9	2.2	4.3	0.8	2.0	3.9
All others	0.5	1.2	2.3	0.6	1.6	3.1	0.7	1.6	3.2
Maintenance	2.4	3.8	5.9	2.0	3.3	5.2	2.7	4.6	7.5
Analysis	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Total conversion costs	4.7	8.0	13.1	4.6	8.4	14.2	5.3	9.5	16.2
Total direct costs	5.2	9.1	15.3	6.5	13.0	23.0	5.9	10.9	18.9
Plant and administrative overheads	2.1	3.1	4.6	1.9	2.7	4.1	2.3	3.5	5.5
Total first-year operating and maintenance costs	7.3	12.2	19.9	8.4	15.7	27.1	8.2	14.4	24.4
Levelized capital charges	8.0	13.6	22.9	9.3	17.0	30.4	9.8	17.8	31.8
Total first-year annual revenue requirements	15.3	25.8	42.8	17.7	32.7	57.5	18.0	32.2	56.2
Levelized annual revenue requirements	21.8	36.6	60.4	25.1	46.5	81.4	25.2	45.1	77.7

TABLE 27. ANNUAL REVENUE REQUIREMENT SUMMARY FOR EXISTING 200-, 500-, AND 1,000-MW POWER UNITS WITH 3.5% SULFUR COAL

	M$								
	Advanced limestone			Davy S-H			Dowa		
	200	500	1,000	200	500	1,000	200	500	1,000
Raw materials									
Limestone	0.4	1.0	1.9	-	-	-	0.4	0.9	1.8
Lime	-	-	-	1.8	4.4	8.7	-	-	-
Adipic acid	0.1	0.2	0.4	-	-	-	-	-	-
Formic acid	-	-	-	0.1	0.2	0.3	-	-	-
Ferric chloride	-	-	-	<0.1	<0.1	0.1	-	-	-
Antifoulant	-	-	-	<0.1	<0.1	<0.1	-	-	-
Aluminum sulfate	-	-	-	-	-	-	0.2	0.5	1.0
Total raw materials	0.5	1.2	2.3	1.9	4.6	9.1	0.6	1.4	2.8
Total labor	1.0	1.2	1.6	1.0	1.2	1.6	1.0	1.2	1.6
Utilities									
Electricity	0.8	1.9	3.6	0.9	2.3	4.5	0.9	2.1	4.0
All others	0.5	1.2	2.4	0.7	1.6	3.1	0.7	1.6	3.1
Maintenance	2.7	5.0	8.2	2.6	4.3	6.8	3.5	6.0	9.9
Analysis	0.1	0.1	0.2	0.1	0.1	0.1	0.1	0.1	0.2
Total conversion costs	5.1	9.4	16.0	5.3	9.5	16.1	6.2	11.0	18.8
Total direct costs	5.6	10.6	18.3	7.2	14.1	25.2	6.8	12.4	21.6
Plant and administrative overheads	2.3	3.8	6.0	2.2	3.3	5.1	2.8	4.4	7.0
Total first-year operating and maintenance costs	7.9	14.4	24.3	9.4	17.4	30.3	9.6	16.8	28.6
Levelized capital charges	9.2	18.2	32.6	12.4	22.8	41.3	13.1	24.1	43.3
Total first-year annual revenue requirements	17.1	32.6	56.9	21.8	40.2	71.6	22.7	40.9	71.9
Levelized annual revenue requirements	22.0	41.5	72.1	27.8	51.2	90.7	28.7	51.4	90.0

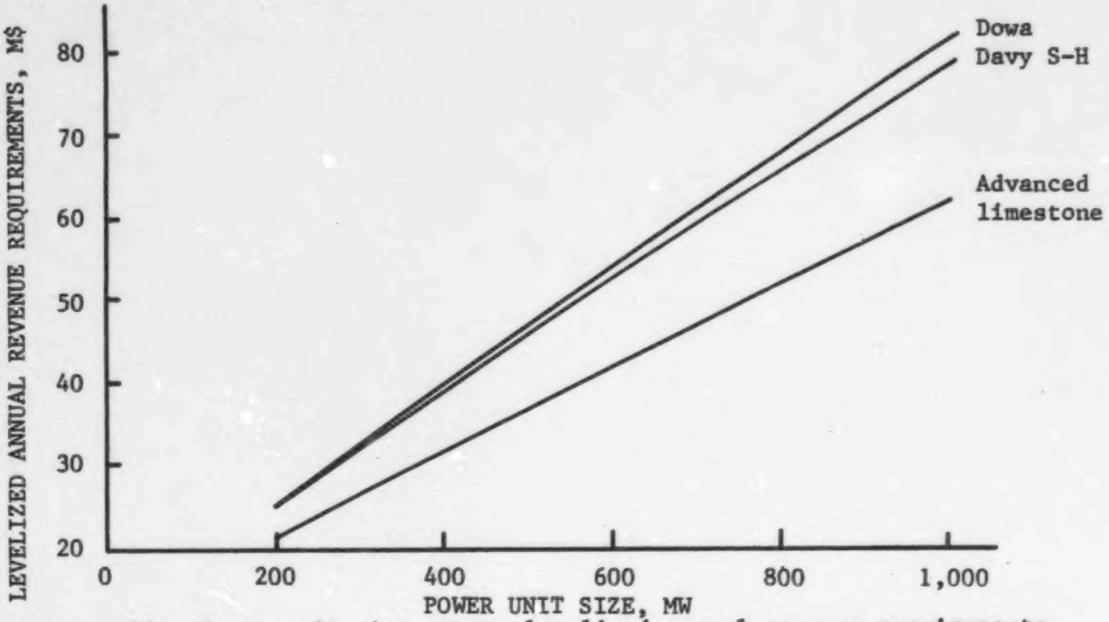


Figure 16. Power unit size versus levelized annual revenue requirements, new units.

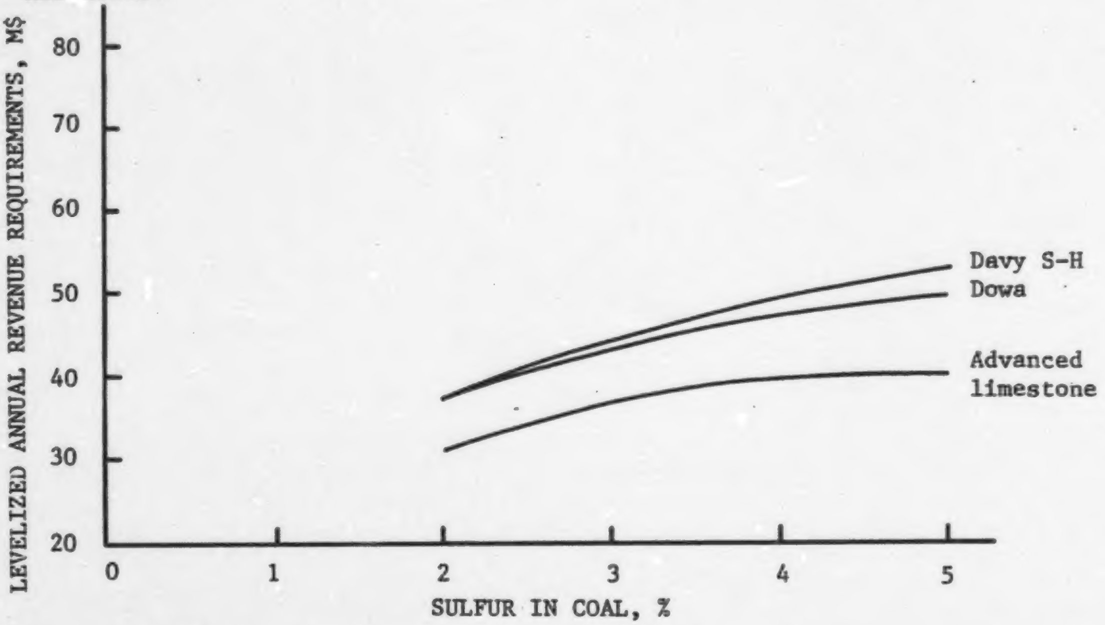


Figure 17. Coal sulfur content versus levelized annual revenue requirements.

cost and capital charges are also highest in every case; but the Davy S-H process has the highest raw material cost in every case. For the 200-MW power unit, the raw material cost difference between the Davy S-H and Dowa processes is insufficient to overcome the maintenance cost and capital charge difference. However, as the power unit size (or coal sulfur level) increases, the raw material cost difference surpasses and overshadows the capital charge difference.

The effect of coal sulfur content on levelized annual revenue requirements is shown in Table 28 and Figure 17. The Davy S-H process has the highest cost, followed in decreasing order by the Dowa and advanced limestone processes. The levelized annual revenue requirements of the Davy S-H process are equivalent to the Dowa process for 2.0% sulfur coal. This trend is due to the relationship of raw material costs, maintenance costs, and capital charges.

As the coal sulfur increases from 2.0% to 3.5%, the levelized annual revenue requirements increase 20% for both the advanced limestone and Dowa processes and 25% for the Davy S-H process. As the coal sulfur increases from 2.0% to 5.0%, the levelized annual revenue requirements increase 30% and 35%, respectively.

Costs related to sulfur removal rates ($/lb SO_2 removed) are reduced by more than half as the coal sulfur increases from 2.0% to 5.0%. This reduction would be greater if it were not for the difference in the controlled outlet emission rates between the 5.0% sulfur coal and the 2.0% and 3.5% sulfur coals; i.e., the 2.0% and 3.5% sulfur coals are limited to an emission level of 0.6 lb SO_2/MBtu while the 5.0% sulfur coal is limited only by a 90% SO_2 removal requirement, which equates to an emission level of 0.82 lb SO_2/MBtu.

DOWA L/G RATIO CASE VARIATION

The Dowa design factors used in this study were approved by UOP based on preliminary evaluation of data from the Dowa process demonstration at the TVA Shawnee test facility near Paducah, Kentucky. However, by mid-1981 with the release of the final test report (83), the conclusions stated that to achieve a SO_2 removal efficiency of 90%, a L/G ratio of approximately 80 gal/kaft^3 was necessary. The preliminary design data had shown a L/G ratio of 45 gal/kaft^3. The preceding design and analysis were based on a L/G ratio of 45 gal/kaft^3 and for this reason, an additional case variation analyzing the effects of the increased L/G ratio was performed.

Capital Investment

The two cost areas that are affected most are the SO_2 absorption and oxidation areas. In the SO_2 absorption area, only the effluent recirculation system is affected. To handle the additional pumping requirements, 700-hp pumps are required. Ancillary equipment such as piping, instrument sizes, and electrical equipment is also increased to handle the additional volume of

TABLE 28. ANNUAL REVENUE REQUIREMENT SUMMARY FOR NEW 500-MW POWER UNITS WITH 2.0%, 3.5%, AND 5.0% SULFUR COAL

	M								
	Advanced limestone			Davy S-H			Dowa		
	2.0	3.5	5.0	2.0	3.5	5.0	2.0	3.5	5.0
Raw materials									
Limestone	0.5	0.9	1.3	-	-	-	0.4	0.9	1.3
Lime	-	-	-	2.2	4.3	6.2	-	-	-
Adipic acid	0.1	0.2	0.3	-	-	-	-	-	-
Formic acid	-	-	-	0.1	0.2	0.2	-	-	-
Ferric chloride	-	-	-	<0.1	<0.1	<0.1	-	-	-
Antifoulant	-	-	-	<0.1	<0.1	<0.1	-	-	-
Aluminum sulfate	-	-	-	-	-	-	0.3	0.5	0.7
Total raw materials	0.6	1.1	1.6	2.3	4.6	6.5	0.7	1.4	2.0
Total labor	1.1	1.2	1.3	1.2	1.2	1.3	1.2	1.2	1.3
Utilities									
Electricity	1.3	1.7	1.8	2.0	2.2	2.3	1.6	2.0	2.2
All others	0.8	1.2	1.3	1.0	1.6	1.6	1.0	1.6	1.6
Maintenance	3.3	3.8	4.0	2.9	3.3	3.4	4.0	4.6	4.7
Analysis	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Total conversion costs	6.6	8.0	8.5	7.2	8.4	8.7	7.9	9.5	9.9
Total direct costs	7.2	9.1	10.1	9.5	13.0	15.2	8.6	10.9	11.9
Plant and administrative overheads	2.7	3.1	3.2	2.5	2.7	2.9	3.1	3.5	3.7
Total first-year operating and maintenance costs	9.9	12.2	13.3	12.0	15.7	18.1	11.7	14.4	15.6
Levelized capital charges	11.7	13.6	14.4	14.7	17.0	17.8	15.1	17.8	19.2
Total first-year annual revenue requirements	21.6	25.8	27.7	26.7	32.7	35.9	26.8	32.2	34.8
Levelized annual revenue requirements	30.4	36.6	39.5	37.3	46.5	52.0	37.2	45.1	48.6

effluent. In the oxidation area, a larger oxidation tank and agitator are required to handle the additional scrubber effluent. The only other adjustments required are for electrical equipment and concrete foundations. Adjustments to other areas are minor. With these adjustments, the SO_2 absorption area and oxidation area costs increase 12.8% and 10.2%, respectively, resulting in a 4.1% increase in the capital investment.

Annual Revenue Requirements

Two components of the total conversion cost, electricity and maintenance, are affected by increasing the L/G ratio from 45 to 80 gal/kaft^3. Due to the increase in horsepower of the effluent recirculation pumps (8 operating at 700 hp each) and agitators (4 operating at 90 hp each), the electricity use increases 19%. Maintenance costs are also increased since the direct process investment increases. With both of these increases taken into account, the total conversion costs increase approximately 6.4%. The levelized annual revenue requirements increased 4.8%.

VARIATIONS IN ECONOMIC FACTORS

To reflect the sensitivity of annual revenue requirements to technical, economic, and geographical variations, major cost elements were varied. The factors evaluated and the ranges of the variations are listed in Table 29.

TABLE 29. ANNUAL REVENUE REQUIREMENT COST ELEMENT VARIATIONS

Cost element	Base value	Range of variations, %
Raw materials[a]	Base case values for raw materials (see Table 10)	150, 200
Electricity	0.037 $/kWh	150, 200
Maintenance	Percentage of total direct investment excluding landfill plus percentage of direct landfill investment	75, 125, 150
Labor cost[b]	FGD labor: 15 $/man-hr Analysis and solids disposal labor: 21 $/man-hr	150, 200

a. All components of raw materials were varied at the same rate; i.e., total raw materials were treated as a single cost element.

b. All labor costs were varied at the same rate. Labor costs include FGD, solids, disposal, and analysis costs; i.e., all labor costs were treated as a single cost element.

The effect of variations in raw materials cost is shown in Figure 18 for different power unit sizes and in Figure 19 for different coal sulfur contents. The Davy S-H process is most sensitive to variations in raw material cost. For example, the differences in levelized annual revenue requirements between the raw material base cost and the 200% base cost for the advanced limestone and Dowa processes are 4%, 6%, and 7% for 200-, 500-, and 1,000-MW power units, respectively, while for the Davy S-H process, the differences are 14%, 18%, and 20%. For the coal sulfur variations, the differences for the 2.0%, 3.5%, and 5.0% coal sulfur variations are 4%, 6%, and 8%, respectively, for the advanced limestone and Dowa processes, and 12%, 18%, and 24% for the Davy S-H process.

The effect of electricity cost variations on the levelized annual revenue requirements is shown in Figure 20 for different power unit sizes and in Figure 21 for different coal sulfur contents. Electricity accounts for 15% to 21% of the operating and maintenance costs. Doubling the cost of electricity affects all cases a maximum of 10% and all processes are affected equally.

Variations in maintenance cost are shown in Figure 22 for different power unit sizes and in Figure 23 for different coal sulfur contents. The advanced limestone and Dowa processes are only slightly more sensitive to variations in maintenance cost than the Davy S-H process. Differences range from a minimum of -8% for the 75% base maintenance cost to a maximum of 17% for the 150% base maintenance cost.

The effect of variations in labor cost is shown in Figure 24 for different power unit sizes and in Figure 25 for different coal sulfur contents. Labor costs are equivalent for similar cases and changes in labor costs thus show the same trend for all variations. For a 200% base labor cost, the increases in levelized annual revenue requirements for all processes are approximately 15%, 10%, and 9% for the 200-, 500-, and 1,000-MW power unit sizes, respectively, and 8% to 12% for all coal sulfur contents.

Capital charges, a direct function of capital investment, are the other elements that have a major effect on the levelized annual revenue requirements. Levelized capital charges, in addition to being a function of interest and depreciation rates, are a function of the discount rate and annual inflation rate. As shown in Table 30 and Figure 26, the effect of various discount and inflation rates is minimal on the capital investment but significant on the levelized annual revenue requirements. For the most extreme economic position (15% discount and 10% inflation rates), the capital investment increases only 6% from the base case (10% discount and 6% inflation rates) while the levelized annual revenue requirements increase 37%. The larger increase in levelized annual revenue requirements occurs because the levelizing factor is determined by both the annual discount rate and the annual inflation rate. Capital investment is only affected by interest rates. Levelizing factors, levelized capital charges, and interest during construction rates for each economic position are also listed in Table 30.

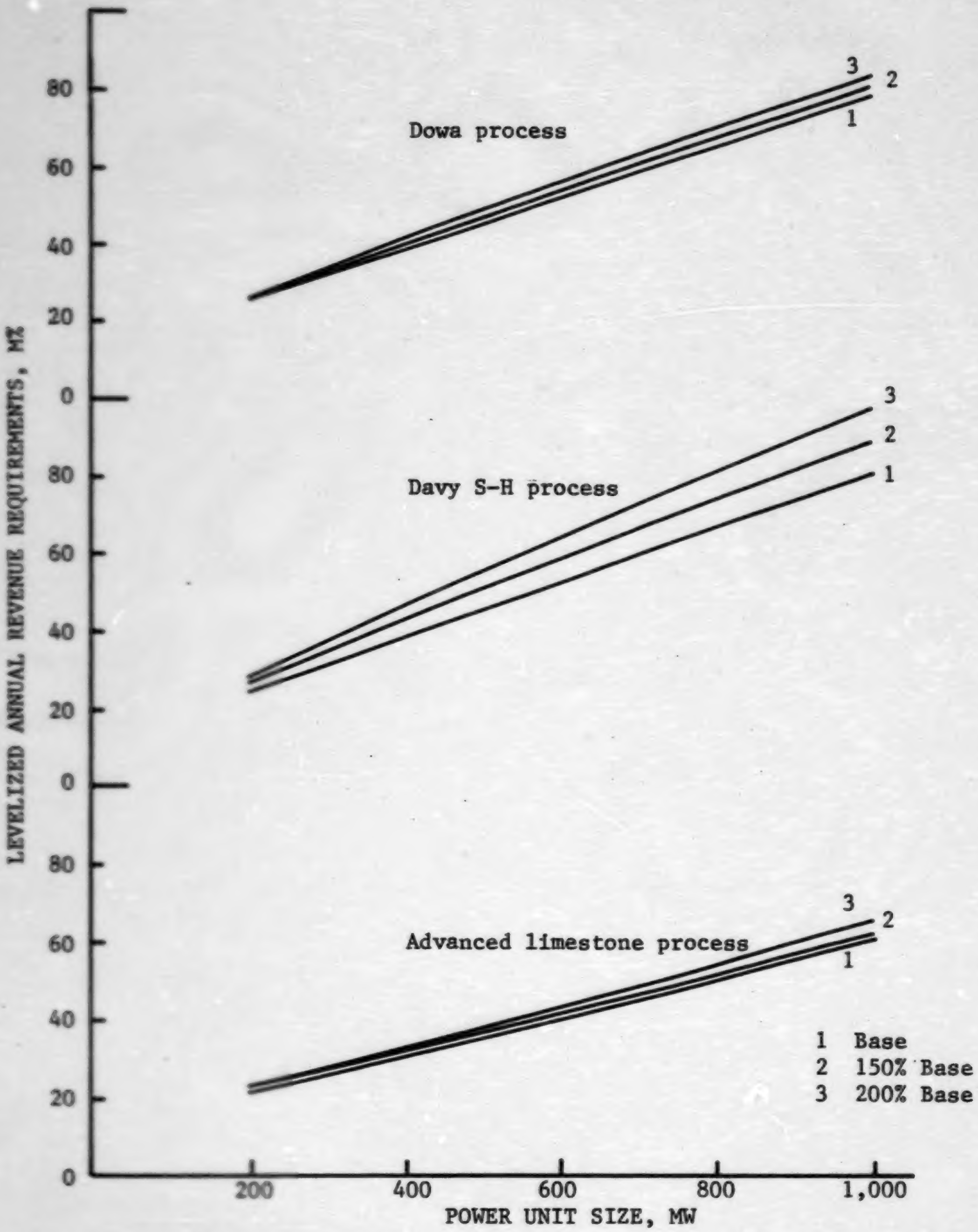


Figure 18. Raw material cost versus levelized annual revenue requirements at different power unit sizes.

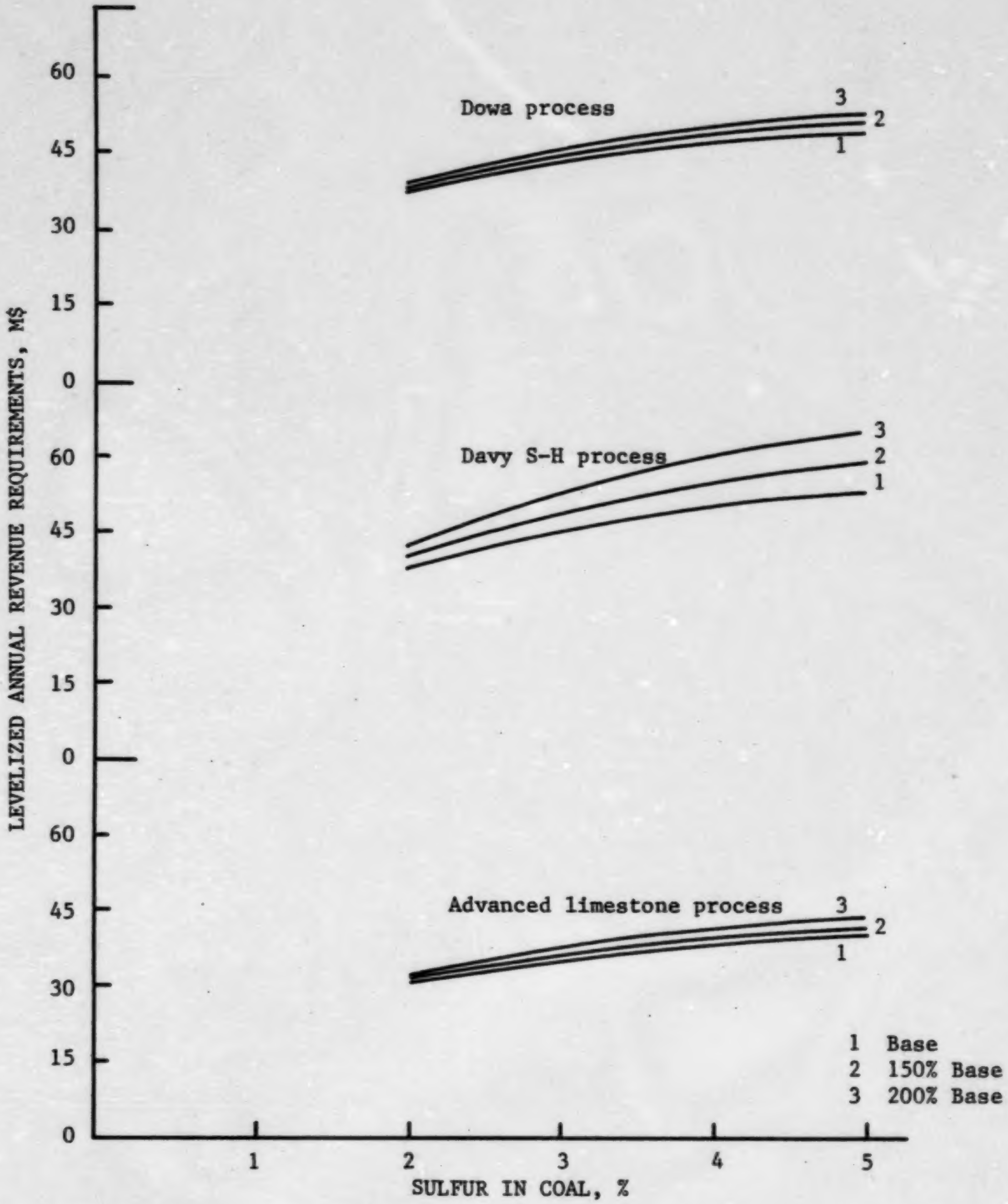


Figure 19. Raw material cost versus levelized annual revenue requirements at different coal sulfur levels.

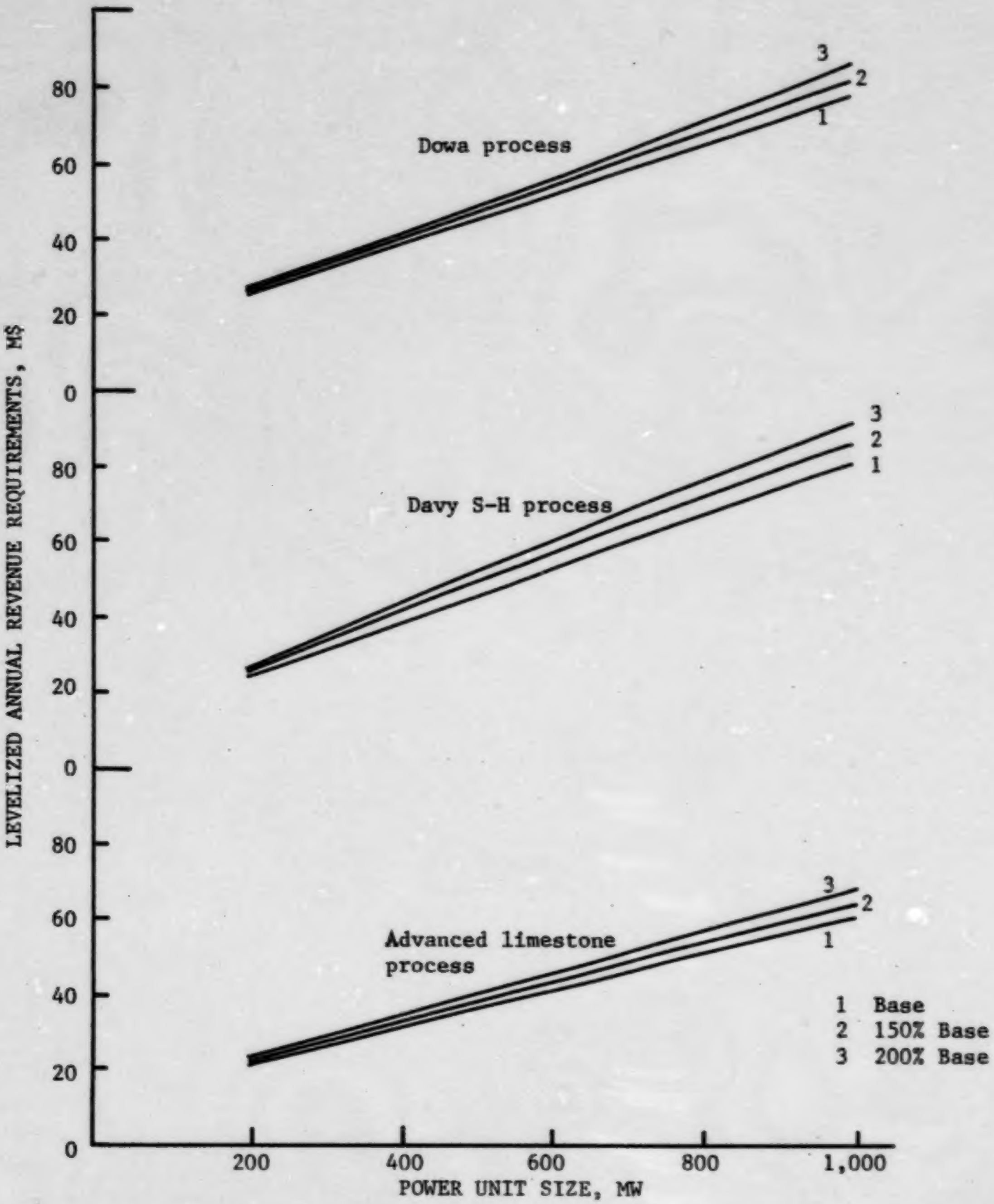


Figure 20. Electricity cost versus levelized annual revenue requirements at different power unit sizes.

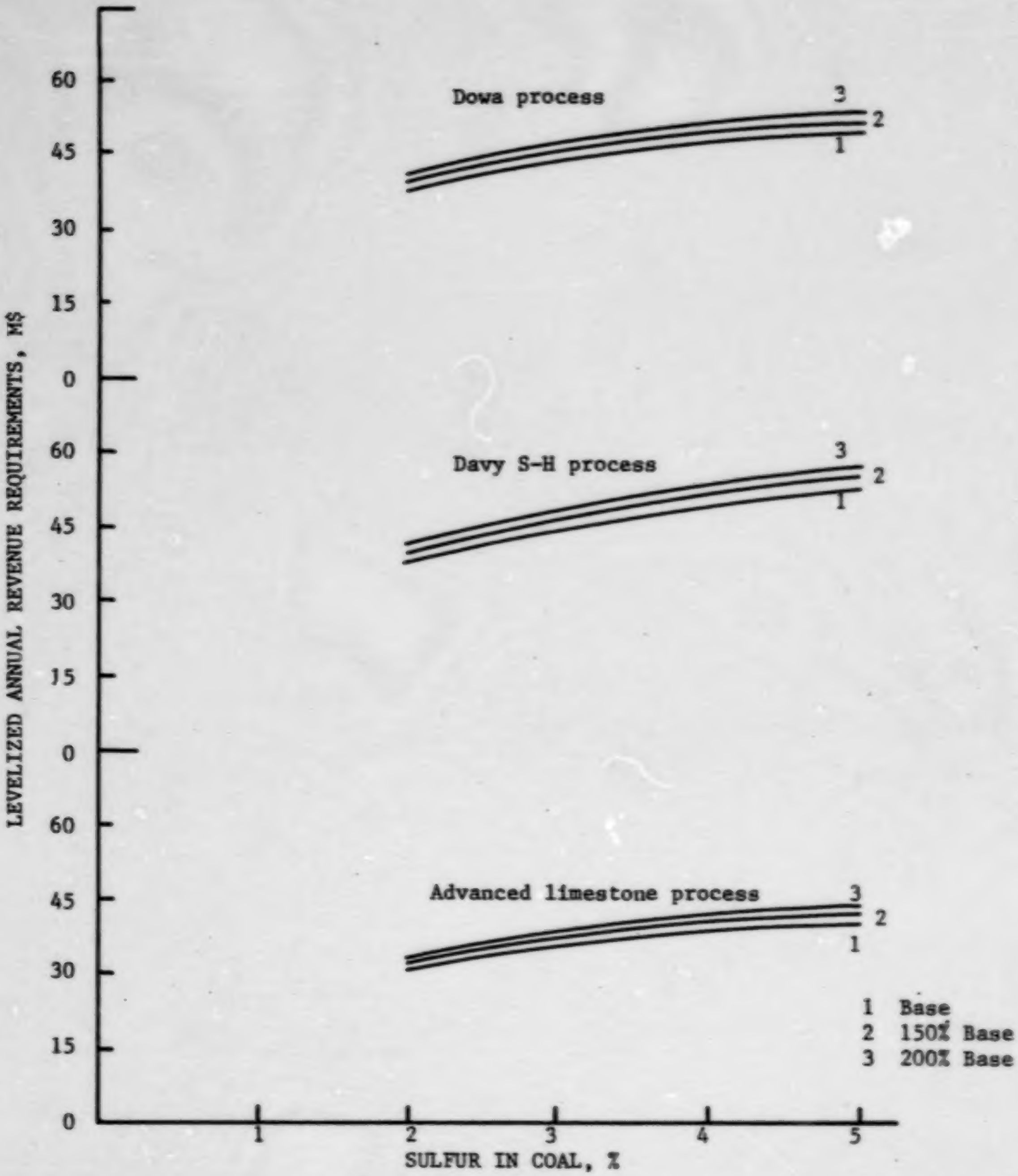


Figure 21. Variations in electricity cost versus levelized annual revenue requirements at different coal sulfur contents.

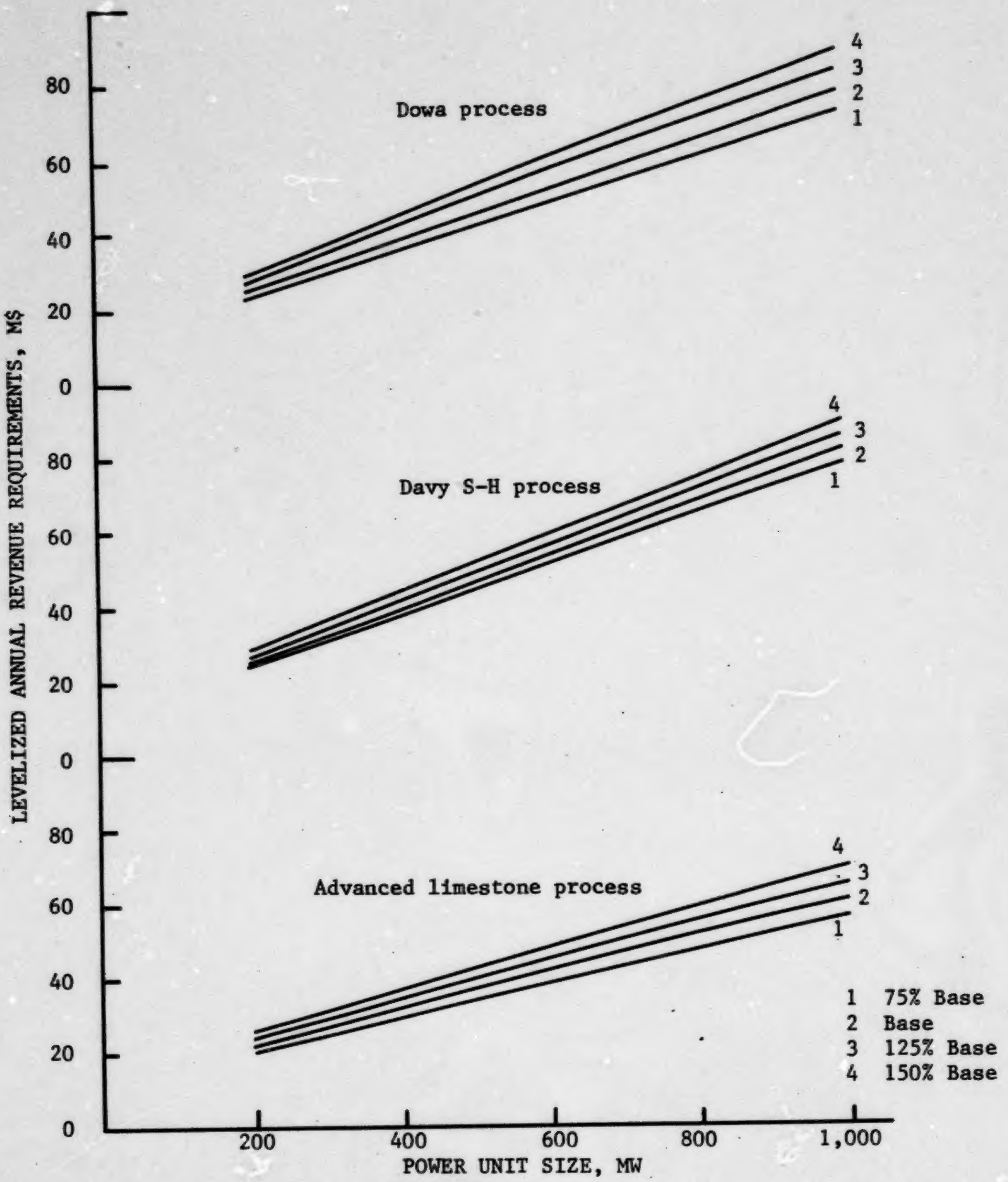


Figure 22. Variations in maintenance cost versus levelized annual revenue requirements at different power unit sizes.

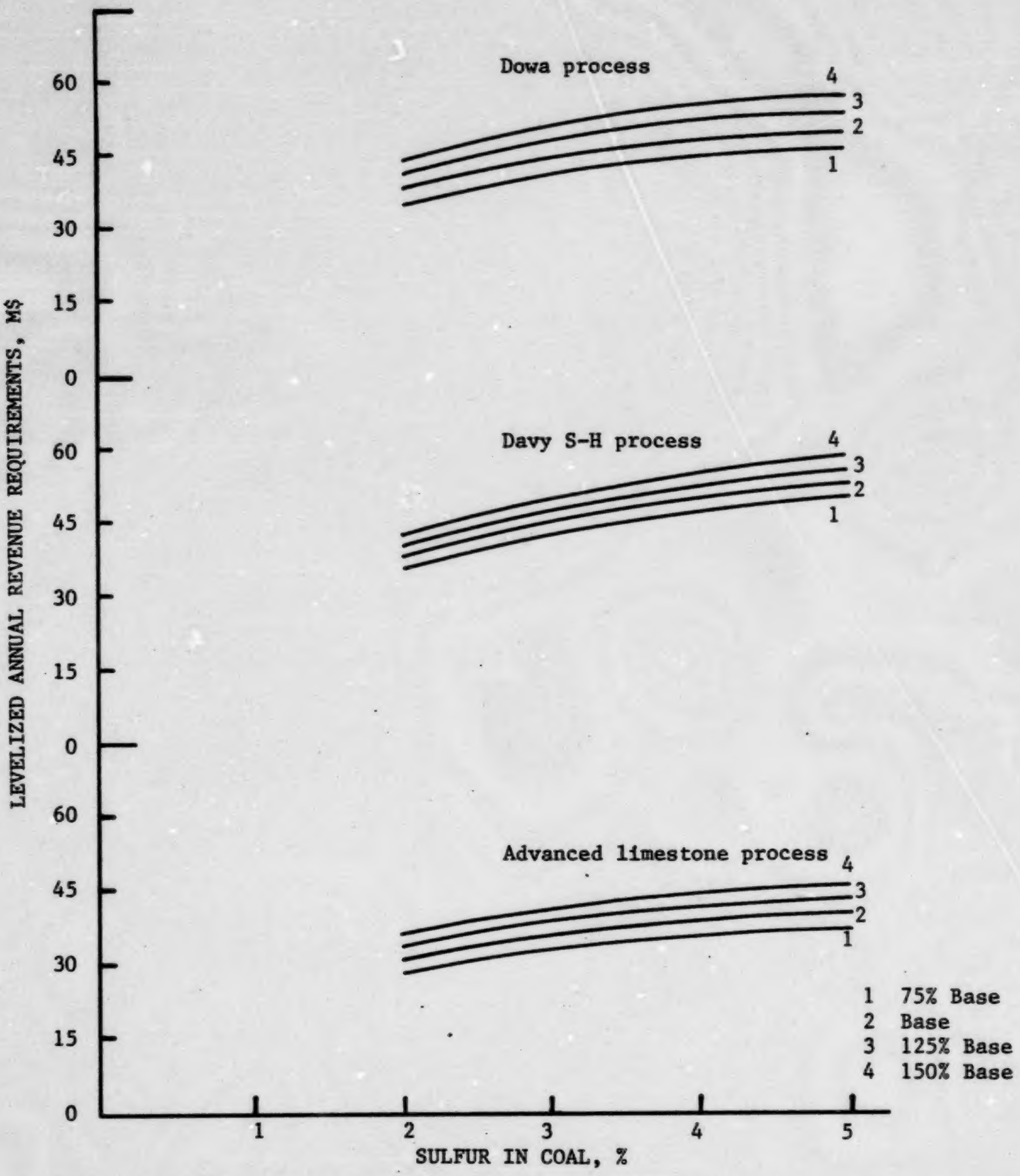


Figure 23. Variations in maintenance cost versus levelized annual revenue requirements at different coal sulfur levels.

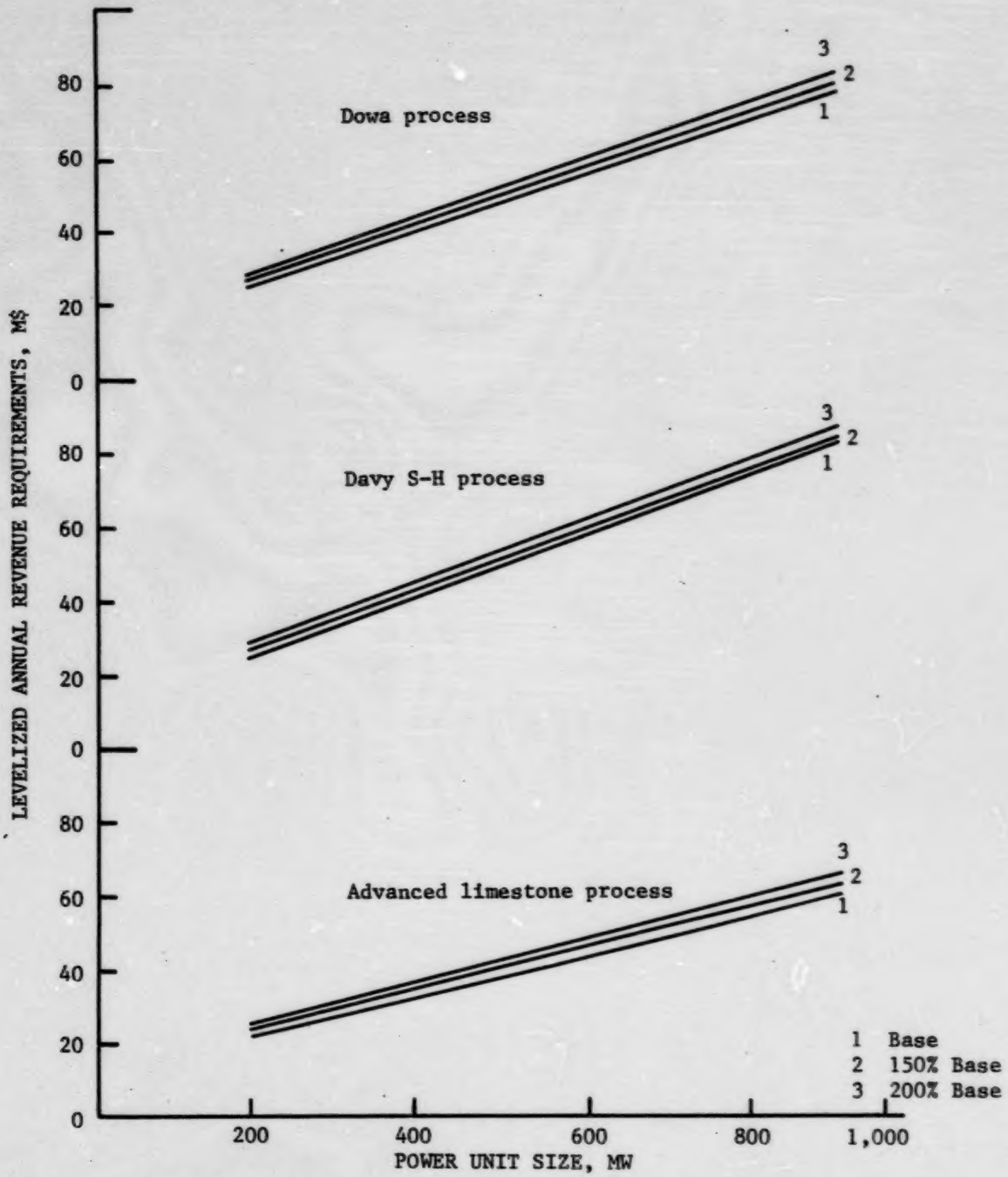


Figure 24. Variations in labor cost versus levelized annual revenue requirements at different power unit sizes.

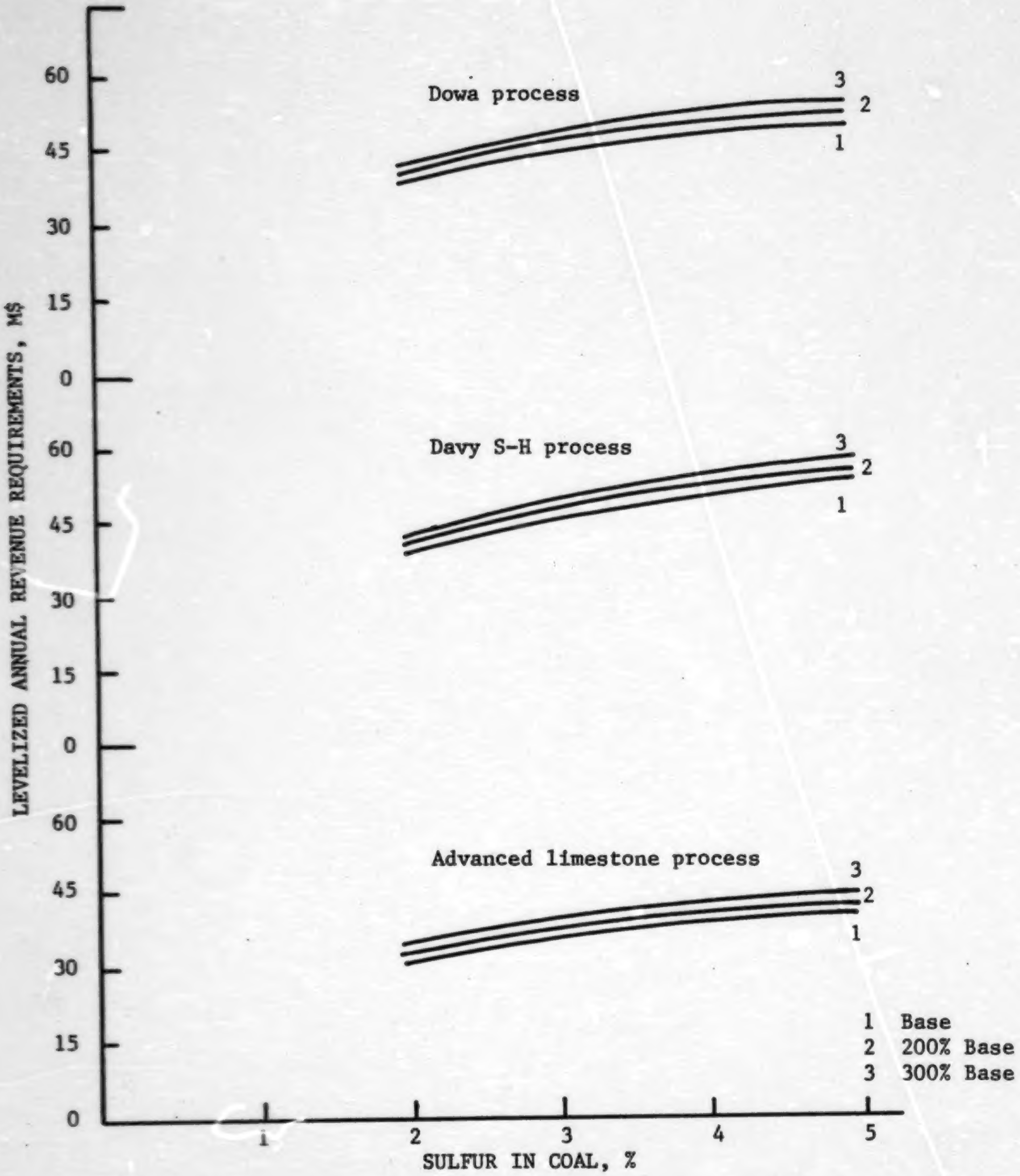


Figure 25. Variations in labor cost versus levelized annual revenue requirements at different coal sulfur levels.

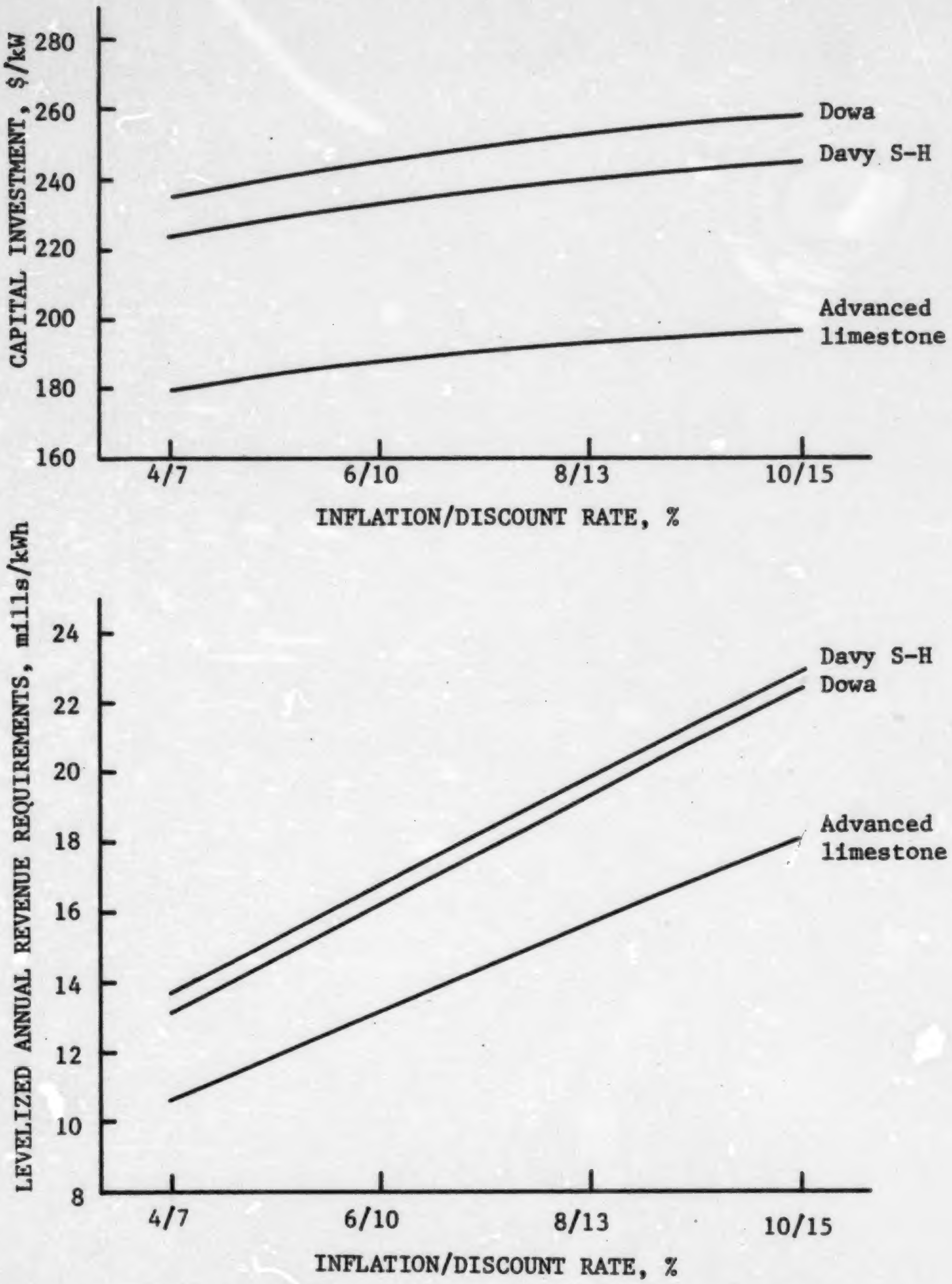


Figure 26. Sensitivity to changes in inflation and discount rate.

TABLE 30. EFFECT OF CAPITAL CHARGES VARIATIONS ON BASE CASE CAPITAL INVESTMENT AND LEVELIZED ANNUAL REVENUE REQUIREMENTS

Discount rate, %: Process Inflation rate, %:	7 4	10 6	13 8	15 10
Advanced limestone				
Total capital investment, M$	89.3	92.8	96.3	98.7
Levelized annual revenue requirements, M$	29.3	36.6	42.3	49.9
Davy S-H				
Total capital investment, M$	111.3	115.5	119.8	122.7
Levelized annual revenue requirements, M$	37.3	46.6	53.8	63.4
Dowa				
Total capital investment, M$	116.8	121.3	125.9	129.0
Levelized annual revenue requirements, M$	35.9	45.1	52.2	61.6
Levelizing factor	1.604	1.886	2.140	2.468
Levelized capital charges, %	10.9	14.7	16.9	20.1
Interest during construction, %	10.8	15.6	20.5	23.8

In summary, under similar conditions only the Davy S-H process is affected by a cost element (raw material) to a greater extent than the other two processes. Electricity, maintenance, and labor cost variations affect each process similarly and therefore, show no strong contrasting trends. Changes in the capital charges related to various economic conditions significantly affect the levelized annual revenue requirements for all processes, but do not change the process relationships since all processes are acted upon by the same factors. Other cost elements which were not explored (diesel fuel, steam, and process water) are minor costs in which variations would not have an appreciable effect on annual revenue requirements.

CONCLUSIONS

The capital investment and annual revenue requirements of the advanced limestone process are about one-fourth and one-tenth lower than those of the Davy S-H and Dowa processes. Much of this difference is the result of a more extensive use of partial bypass for the advanced limestone process, made possible by the use of adipic acid, which allows scrubbing at 95% removal efficiency as compared with 90% for the other processes. This reduces the capital investment and operating costs such as electricity and reheat steam requirements. The capital investments and annual revenue requirements of the Davy S-H and Dowa processes differ no more than a few percent; the Davy S-H process is slightly lower in capital investment and slightly higher in annual revenue requirements than the Dowa process. The relative differences in costs of the processes do not change appreciably for other conditions of power unit size and age and coal sulfur content.

In capital investment, the Davy S-H process has the lowest feed preparation costs (because lime is used) and the highest gas-handling costs (because of a high pressure drop and more extensive use of stainless steel); the advanced limestone process has the highest SO_2 absorption costs (because of the high L/G ratio and the additional recirculation tank) and the lowest gas-handling costs (because of the partial bypass); and the oxidation costs are higher for both the Davy S-H and Dowa processes (because of a special design in the former and a stainless steel sparger in the latter).

The only significant difference in the distribution of annual revenue requirements by major cost components is in raw material costs. The Davy S-H process raw material costs are four times those of the advanced limestone process because lime is used. Those of the Dowa process are 1.3 times those of the advanced limestone process because of makeup aluminum sulfate costs. Other raw material costs (adipic acid, formic acid, and ferric chloride) are insignificant.

An increase in the L/G ratio for the Dowa process from 45 to 80 gal/kaft^3 increases the capital investment and annual revenue requirements by 4% to 5%. The difference, if a higher L/G ratio was necessary, would not have a material effect on the cost relationships of the processes. Similarly, the chloride purge used in the Dowa process, but not in the other two processes, had no significant effect on the costs.

Landfill capital investments and annual revenue requirements are a minor cost element for all of the processes.

REFERENCES

1. S. V. Tomlinson, F. M. Kennedy, F. A. Sudhoff, and R. L. Torstrick, 1979, Definitive SO_x Control Process Evaluations: Limestone, Double Alkali, and Citrate FGD Processes, EPA-600/7-79-177, U.S. Environmental Protection Agency, Washington, D.C.

2. K. D. Anderson, J. W. Barrier, W. E. O'Brien, and S. V. Tomlinson, 1980, Definitive SO_x Control Process Evaluations: Limestone, Lime, and Magnesia Processes, EPA-600/7-80-001, U.S. Environmental Protection Agency, Washington, D.C.

3. J. R. Byrd, K. D. Anderson, S. V. Tomlinson, and R. L. Torstrick, 1981, Definitive SO_x Control Process Evaluations: Aqueous Carbonate and Wellman-Lord (Acid, Allied Chemical, and Resox) FGD Technologies, EPA-600/7-81-099, U.S. Environmental Protection Agency, Washington, D.C.

4. T. A. Burnett, C. D. Stephenson, F. A. Sudhoff, and J. D. Veitch, 1983, Economic Evaluation of Advanced Limestone and Lime Flue Gas Desulfurization Processes, EPA-600/7-83-029, U.S. Environmental Protection Agency, Washington, D.C.

5. Title 40, Code of Federal Regulations, Part 60.

6. H. F. Johnstone, 1931, Metallic Ions as Catalysts for the Removal of Sulfur Dioxide from Boiler Furnace Gases, Industrial and Engineering Chemistry, Vol. 23, No. 5, pp. 559-561.

7. R. L. Copson and J. W. Payne, 1933, Recovery of Sulfur Dioxide as Dilute Sulfuric Acid, Industrial and Engineering Chemistry, Vol. 25, No. 8, pp. 909-916.

8. G. Tarbutton, J. C. Driskell, T. M. Jones, F. J. Gray, and C. M. Smith, 1957, Recovery of Sulfur Dioxide from Flue Gases, Industrial and Engineering Chemistry, Vol. 49, No. 3, pp. 392-395.

9. G. Nonhebel, 1936, A Commercial Plant for Removal of Smoke and Oxides of Sulphur from Flue Gases, Transactions of the Faraday Society, Vol. XXXII, pp. 1291-1297.

10. L. C. McCabe, 1955, Atmospheric Pollution, Industrial and Engineering Chemistry, Vol. 47, No. 4, pp. 99A-101A.

Preceding page blank

11. R. Holighaus, 1978, Desulfurization Technology in the Federal Republic of Germany, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, Vol. 1, EPA-600/7-78-058a, U.S. Environmental Protection Agency, Washington, D.C., pp. 80-95.

12. J. Ando, 1974, Utilizing and Disposing of Sulfur Products From Flue Gas Desulfurization Processes in Japan, In: Proceedings: Symposium on Flue Gas Desulfurization - Atlanta, November 1974, Vol. II, EPA-650/2-74-126-b, U.S. Environmental Protection Agency, Washington, D.C., pp. 955-968.

13. M. Smith, M. Melia, N. Gregory, and K. Scalf (PEDCo Environmental, Inc.), 1981, EPA Utility FGD Survey: October-December 1980, Vol. 1, EPA-600/7-81-102a, U.S. Environmental Protection Agency, Washington, D.C.

14. R. G. Knight, E. H. Rothfuss, and K. D. Yard, (Michael Baker Jr., Inc.), 1980, FGD Sludge Disposal Manual, Second Edition, CS-1515, Electric Power Research Institute, Palo Alto, CA.

15. R. J. Kruger, 1978, Experience with Limestone Scrubbing, Sherburne County Generating Plant, Northern States Power Company, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, Vol. I, EPA-600/7-78-058a, U.S. Environmental Protection Agency, Washington, D.C., pp. 292-319.

16. L. M. Pruce, 1981, Why So Few Regenerative Scrubbers?, Power, Vol. 125, No. 6, pp. 73-76.

17. G. T. Rochelle and C. J. King, 1977, Process Alternates for Stack Gas Desulfurization by Throwaway Scrubbing, In: Proceedings: The Second Pacific Chemical Engineering Congress (Pachec '77), Vol. I, the American Institute of Chemical Engineers, New York, NY, pp. 261-267.

18. A. G. Clarke, 1981, Electrolyte Solution Theory and the Oxidation Rate of Sulphur Dioxide in Water, Atmospheric Environment, Vol. 15, No. 9, pp. 1591-1595.

19. R. H. Borgwardt, 1976, IERL-RTP Scrubber Studies Related to Forced Oxidation, In: Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, March 1976, Vol. 1, EPA-600/2-76-136a, U.S. Environmental Protection Agency, Washington, D.C., pp. 117-143.

20. R. H. Borgwardt, 1977, Sludge Oxidation in Limestone FGD Scrubbers, EPA-600/7-77-061, U.S. Environmental Protection Agency, Washington, D.C.

21. M. Noguchi, 1974, Status Report on Chiyoda Thoroughbred 101 Process, In: Proceedings: Symposium on Flue Gas Desulfurization - Atlanta, November 1974, Vol. II, EPA-650/2-74-126-b, U.S. Environmental Protection Agency, Washington, D.C.

22. M. Harai, M. Atsukawa, A. Tatani, and K. Kondo, 1976, *MHI Flue Gas Desulfurization Systems Applied to Several Emission Sources*, *In*: Proceedings: Second International Lime/Limestone Wet-Scrubbing Symposium, November 1971, New Orleans, Vol. 1, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C., pp. 249-268.

23. R. J. Gleason, 1977, *Improved Flue Gas Desulfurization Process with Oxidation*, *In*: Proceedings: The Second Pacific Chemical Engineering Congress (Pachec '77), Vol. I, American Institute of Chemical Engineers, New York, NY, pp. 371-377.

24. H. N. Head and S.-C. Wang, 1979, *EPA Alkali Scrubbing Test Facility: Advanced Program, Fourth Progress Report*, 2 volumes, EPA-600/7-79-244a and -244b, U.S. Environmental Protection Agency, Washington, D.C.

25. S. Saito, T. Morita, and S. Suzuki, 1976, *Kureha Flue Gas Desulfurization "Sodium Acetate-Gypsum Process,"* *In*: Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, March 1976, Vol. I, EPA-600/2-76-136a, U.S. Environmental Protection Agency, Washington, D.C., pp. 515-543.

26. G. P. Behrens and O. W. Hargrove, 1980, *Evaluation of Chiyoda Thoroughbred 121 FGD Process and Gypsum Stacking*, Vols. 1 and 2, CS-1579, Electric Power Research Institute, Palo Alto, CA.

27. G. T. Rochelle, 1977, *Process Synthesis and Innovation in Flue Gas Desulfurization*, pH.D. Dissertation, University of California at Berkeley.

28. W. E. Corbett, O. W. Hargrove, and R. S. Merrill, (Radian Corp.), 1977, *A Summary of the Effects of Important Chemical Variables Upon the Performance of Lime/Limestone Wet Scrubbing Systems*, FP-639 Electric Power Research Institute, Palo Alto, CA.

29. P. S. Lowell, D. M. Ottmers, Jr., K. Schwitzgebel, T. I. Strange, and D. W. DeBerry, (Radian Corp.), 1970, *A Theoretical Description of the Limestone Injection - Wet Scrubbing Process*, Final report to the National Air Pollution Control Administration, NTIS PB 193029.

30. H. N. Head, 1977, *EPA Alkali Scrubbing Test Facility: Advanced Program, Third Progress Report*, EPA-600/7-77-105, U.S. Environmental Protection Agency, Washington, D.C.

31. H. N. Head, 1976, *EPA Alkali Scrubbing Test Facility: Advanced Program, Second Progress Report*, EPA-600/7-76-008, U.S. Environmental Protection Agency, Washington, D.C.

32. E. Hayon, A. Treinin, and J. Wilf, 1972, *Electric Spectra, Photochemistry, and Autoxidation Mechanism of the Sulfite-Bisulfite-Pyrosulfite Systems. The SO_2^-, SO_3^-, SO_4^-, and SO_5^- Radicals*, Journal of the American Chemical Society, Vol. 94, No. 1, pp. 47-57.

33. T. V. Larson, N. R. Horike, and H. Harrison, 1978, *Oxidation of Sulfur Dioxide by Oxygen and Ozone in Aqueous Solution: A Kinetic Study with Significance to Atmospheric Rate Processes*, Atmospheric Environment, Vol. 12, pp. 1597-1611.

34. H. N. Alyea and H. L. Backstrom, 1929, *The Inhibitive Action of Alcohols on the Oxidation of Sodium Sulfite*, Journal of The American Chemical Society, Vol. 51, pp. 90-109.

35. C. H. Barron and H. A. O'Hern, 1966, *Reaction Kinetics of Sodium Sulfite Oxidation by the Rapid-Mixing Method*, Chemical Engineering Science, Vol. 21, pp. 397-404.

36. V. Linek and J. Mayrhoferova, 1970, *The Kinetics of Oxidation of Aqueous Sodium Sulphite Solution*, Chemical Engineering Science, Vol. 25, pp. 787-800.

37. K.J.A. de Wall and J. C. Okeson, 1966, *The Oxidation of Aqueous Sodium Sulphite Solutions*, Chemical Engineering Science, Vol. 21, pp. 559-572.

38. E. C. Fuller and R. H. Crist, 1941, *The Rate of Oxidation of Sulfite Ions by Oxygen*, Journal of The American Chemical Society, Vol. 63, pp. 1644-1650.

39. H. S. Rosenberg and H. M. Grotta, 1980, *NO_x Influence on Sulfite Oxidation and Scaling in Lime/Limestone Flue Gas Desulfurization (FGD) Systems*, Environmental Science & Technology, Vol. 14, No. 4, pp. 470-472.

40. S. E. Sadek, D. A. Nawrocki, E. E. Sterbis, and J. E. Vivian, 1977, *Absorption with Reaction of Sulfur Dioxide and Chlorine, Industrial and Engineering Chemistry*, Fundamentals, Vol. 16, No. 1, pp. 36-39. See also, S. Whitaker, *A Comment on the Chlorine-Sulfur Dioxide Reaction in Aqueous Solution*, Vol. 16, No. 3, p. 391.

41. Y. Yamamichi and J.-i. Nagao, 1976, *Removal of Sulfur Oxide from Waste Gas*, U.S. Patent 3,943,230.

42. D. D. Clasen and H. Idemura, 1978, *Limestone/Gypsum Jet Bubbling Scrubbing System*, *In*: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, Vol. II, EPA-600/7-78-058b, U.S. Environmental Protection Agency, Washington, D.C., pp. 837-856.

43. M. G. Manvelyan, G. O. Grigoryan, S. A. Gazaryan, G. S. Papyan, S. S. Karakhanyan, and G. A. Arutyunyan, 1961, *Effect of Inhibitors on Oxidation of Magnesium Sulfite by Atmospheric Oxygen in Presence of Traces of Nitrogen Oxides*, Journal of Applied Chemistry of the USSR, Vol. 34, pp. 896-898.

44. J. L. Hudson, 1980, *Sulfur Dioxide Oxidation in Scrubber Systems*, EPA-600/7-80-083, U.S. Environmental Protection Agency, Washington, D.C.

45. V. Linek and V. Vacek, 1981, Chemical Engineering Use of Catalyzed Sulfite Oxidation Kinetics for the Determination of Mass Transfer Characteristics of Gas-Liquid Contactors, Chemical Engineering Science, Vol. 36, No. 11, pp. 1747-1768.

46. P. V. Danckwerts, 1970, Gas-Liquid Reactions, McGraw-Hill, New York, NY.

47. J. L. Hudson, J. Erwin, and N. M. Catipovic, 1979, Kinetics of Sulfur Dioxide Oxidation in Aqueous Solution, EPA-600/7-79-030, U.S. Environmental Protection Agency, Washington, D.C.

48. S. Beilke and G. Gravenhorst, 1978, Heterogeneous SO_2- Oxidation in the Droplet Phase, Atmospheric Environment, Vol. 12, pp. 231-239.

49. P. Brimblecombe and D. J. Spedding, 1974, The Catalytic Oxidation of Micromolar Aqueous Sulphur Dioxide - I, Atmospheric Environment, Vol. 8, pp. 937-945.

50. J. E. Freiberg and S. E. Schwartz, 1981, Oxidation of SO_2 in Aqueous Droplets: Mass Transport Limitation in Laboratory Studies and the Ambient Atmosphere, Atmospheric Environment, Vol. 15, No. 7, pp. 1145-1154.

51. W. A. Cronkright, Jr., and B. G. Mandelik, 1975, Removing Sulfur-Containing Gases from Waste Gas, U.S. Patent 3,883,639.

52. W. A. Cronkright and W. J. Leddy, 1976, Improving Mass Transfer Characteristics of Limestone Slurries by Use of Magnesium Sulfate, Environmental Science & Technology Vol. 10, No. 6, pp. 569-572.

53. R. L. Nelson and R. D. O'Hara, 1977, Operating Experiences at the Phillips and Elrama Flue Gas Desulfurization Facilities, In: Proceedings: The Second Pacific Chemical Engineering Congress (Pachec '77), Vol. I, pp. 308-315.

54. D. H. Stowe, D. S. Henzel, D. C. Hoffman, 1980, The FGD Dilemma; Lime, Limestone, or Thiosorbic Lime, In: Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, NV, March 1979; Vol. II, EPA-600/7-79-167b, U.S. Environmental Protection Agency, Washington, D.C., pp. 959-988.

55. C. S. Chang and G. T. Rochelle, 1981, Effect of Organic Acid Additives on SO_2 Absorption into $CaO/CaCO_3$ Slurries, In: Transport with Chemical Reactions, AIChE Symposium Series, P. Stroeve and W. J. Ward, eds., No. 202, Vol. 77, The American Institute of Chemical Engineers, New York, NY, pp. 78-86.

56. J. D. Hatfield and J. M. Potts, 1972, Removal of Sulfur Dioxide from Stack Gases by Scrubbing with Limestone Slurry: Use of Organic Acids, In: Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, Vol. II, APTD-1161, U.S. Environmental Protection Agency, Research Triangle Park, NC., pp. 263-283.

57. H. N. Head, S. C. Wang, D. T. Rabb, R. H. Borgwardt, J. E. Williams, and M. A. Maxwell, 1979, Recent Results from EPA's Lime/Limestone Scrubbing Programs - Adipic Acid as a Scrubber Additive, In: Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, NV, March 1979; Vol. I, EPA-600/7-79-167a, U.S. Environmental Protection Agency, Washington, D.C., pp. 342-385.

58. D. A. Burbank, S. C. Wang, R. R. McKinsey, and J. E. Williams, 1980, Test Results on Adipic Acid-Enhanced Limestone Scrubbing at the EPA Shawnee Test Facility - Third Report, In: Proceedings: Symposium on Flue Gas Desulfurization - Houston, TX, October 1980, Vol. I, EPA-600/9-81-019a, U.S. Environmental Protection Agency, Washington, D.C., pp. 233-280.

59. S. C. Wang and D. A. Burbank, (Bechtel National, Inc.), 1981, Adipic Acid Enhanced Lime/Limestone Test Results at the EPA Alkali Scrubbing Test Facility, preprint, presented at the EPA Industry Briefing on Adipic Acid Enhanced Limestone Flue Gas Desulfurization Process, Springfield, MO, July 15, 1981.

60. J.C.S. Chang and J. D. Mobley, 1981, Overview of EPA's Testing of the Adipic Acid Enhanced Limestone FGD Process, preprint, presented at the EPA Industrial Briefing on Adipic Acid Enhanced Limestone Flue Gas Desulfurization Process, Springfield, MO, July 15, 1981.

61. O. W. Hargrove, J. D. Colley, and G. E. Brown, 1981, Status of Adipic Acid Enhanced FGD System Demonstration at Springfield City Utilities Southwest Power Plant, preprint, presented at the EPA Industrial Briefing on Adipic Acid Enhanced Limestone Flue Gas Desulfurization Process, Springfield, MO, July 15, 1981.

62. R. W. Gerstle and D. S. Henzel, 1981, Evaluation of the Adipic Acid Enhanced Limestone Flue Gas Desulfurization Process on an Industrial Boiler, preprint, presented at the EPA Industrial Briefing on Adipic Acid Enhanced Limestone Flue Gas Desulfurization Process, Springfield, MO, July 15, 1981.

63. F. B. Meserole, D. L. Lewis, and F. T. Kurzawa, 1980, Further Study of Adipic Acid Degradation in FGD Scrubbers, EPA-600/PS-7-80-152, Project Summary, U.S. Environmental Protection Agency, Washington, D.C.

64. J.C.S. Chang, 1981, Effect of Scrubbing Operating Conditions on Adipic Acid Degradation, EPA-600/7-81-017 (NTIS PB 81-163305), U.S. Environmental Protection Agency, Washington, D.C.

65. A.E.Z. Wissa (Ardaman & Associates, Inc.), 1978, Environmental Engineering of Gypsum Stacking, preprint, presented at the 85th National Meeting of the American Institute of Chemical Engineers, Philadelphia, PA, June 1978.

66. J. E. Garlanger and T. S. Ingra, (Ardaman & Associates, Inc.), 1980, Evaluation of Chiyoda Thoroughbred 121 FGD Process and Gypsum Stacking, Volume 3: Testing the Feasibility of Stacking FGD Gypsum, CS-1579 Electric Power Research Institute, Palo Alto, CA. See also, Vol. 3 Addendum, 1981.

67. D. A. Burbank and S. C. Wang, 1980, EPA Alkali Scrubbing Test Facility: Advanced Program - Final Report (October 1974-June 1978), EPA-600/7-80-115, U.S. Environmental Protection Agency, Washington, D.C.

68. D. W. Hupe, 1981, Monitoring the Fixed FGD Sludge Landfill, Conesville, Ohio, Phase II, CS-1984, Electric Power Research Institute, Palo Alto, CA.

69. C. R. Beek, 1979, Sulfur Emission: Control Technology & Waste Management, EPA Decision Series, EPA-600/9-79-019, U.S. Environmental Protection Agency, Washington, D.C.

70. S. G. Vick, 1977, Rehabilitation of a Gypsum Tailings Embankment, In: Proceedings of the Conference on Geotechnical Practices for Disposal of Solid Waste Materials, American Society of Civil Engineers, New York, NY, pp. 697-714.

71. R. B. Fling, P. R. Hurt, J. Rossoff, and J. R. Witz, 1980, Disposal of Flue Gas Cleaning Wastes: EPA Shawnee Field Evaluation - Third Annual Report, EPA-600/7-80-011, U.S. Environmental Protection Agency, Washington, D.C.

72. R. H. Borgwardt, 1978, Effect of Forced Oxidation on Limestone/SO_2 Scrubber Performance, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, Vol. I, EPA-600/7-78-058a, U.S. Environmental Protection Agency, Washington, D.C., pp. 205-228.

73. A. L. Pollock, A. Berndl, and M. Esche, 1981, The Davy S-H Process, sales literature, Davy McKee Corp., Lakeland, FL.

74. N. Haug, G. Oelert, and G. Weisser, 1977, NATO CCMS Flue Gas Desulfurization (FGD) Study, Battelle-Institut e.v., Frankfort am Main, FRG.

75. J. Glamser, 1980, personal communication, Davy Corp., Ltd., Houston, TX.

76. M. Esche, 1978, Saarberg-Holter FGD Process, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, Vol. 1, EPA-600/7-78-058a, U.S. Environmental Protection Agency, Washington, D.C., pp. 321-348.

77. F. Hofmann, H. Igelbuscher, and G. Scholl, 1977, Project Study for Flue Gas Desulfurization Unit for the Weiher III Power Plant (Summary), Research Report T 78-28, Saarbergwerke AG, Saarbrucken, FRG (in German). Translation by Ralph McElrog Co., Austin, TX.

78. J. H. Glamser, K. L. Rock, and M. Esche, 1979, A Lime Based FGD Process Utilizing Forced Oxidation to Produce Stable Gypsum By-product, Combustion, Vol. 51, No. 1, pp. 22-26.

79. D. R. Kirkby, 1979, Flue-Gas Desulfurization Produces Salable Gypsum, Chemical Engineering, Vol. 86, No. 1, pp. 56-57.

80. J. Ando, 1974, Status of Flue Gas Desulfurization Technology in Japan, In: Proceedings: Symposium on Flue Gas Desulfurization - Atlanta, November 1974, Vol. 1, EPA-650/2-74-126a, U.S. Environmental Protection Agency, Washington, D.C., pp. 125-148.

81. J. Ando, 1979, SO_2 and NO_x Removal Technology in Japan, In: Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, NV, U.S. Environmental Protection Agency, Washington, D.C., pp. 418-449.

82. S. B. Jackson, C. E. Dene, and D. B. Smith, 1981, Dowa Process Tests, Shawnee Test Facility, In: Proceedings: Symposium on Flue Gas Desulfurization - Houston, TX, October 1980; Vol. 1, EPA-600/9-81-019a, U.S. Environmental Protection Agency, Washington, D.C., pp. 311-325.

83. E. J. Puschaver, J. D. Veitch, and S. B. Jackson, Dowa Process Demonstration Shawnee Test Facility Final Report, in press, Electric Power Research Institute, Palo Alto, CA.

84. M. Katz and R. J. Cole, 1950, Recovery of Sulfur Compounds from Atmospheric Contaminants, Industrial and Engineering Chemistry, Vol. 42, No. 11, pp. 2258-2269.

85. H. F. Johnstone, H. J. Read, and H. C. Blankmeyer, 1938, Recovery of Sulfur Dioxide from Waste Gases, Industrial and Engineering Chemistry, Vol. 30, No. 1, pp. 101-109.

86. Y. Yamamichi and J. Nagao, 1976, The Dowa's Basic Aluminum Sulfate-Gypsum Flue Gas Desulfurization Process, In: Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, March 1976, Vol. II, EPA-600/2-76-136b, U.S. Environmental Protection Agency, Washington, D.C., pp. 833-842.

87. F. A. Cotton and G. Wilkinson, 1966, Advanced Inorganic Chemistry, Interscience Publishers, New York, NY.

88. J. Aveston, 1965, Hydrolysis of the Aluminum Ion: Ultracentrifugation and Acidity Measurements, Journal of the Chemical Society, 1965, pp. 4438-4443.

89. J. L. Henry and G. B. King, 1949, The System Aluminum Sulfate-Sulfuric Acid - Water at 60°, Journal of the American Chemical Society, Vol. 71, pp. 1142-1144.

90. J. Ando, 1978, Status of SO_2 and NO_x Removal Systems in Japan, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, Vol. 1, EPA-600/7-78-058a, U.S. Environmental Protection Agency, Washington, D.C., pp. 59-79.

91. J. A. Cavallaro, M. J. Johnson, and A. W. Deubrouck, 1976, Sulfur Reduction Potential of the Coals of the United States, Bureau of Mines Report of Investigation RI 8118, U.S. Bureau of Mines, Washington, D.C.

92. J. W. Hamersima and M. L. Kraft, 1975, Applicability of the Meyers Process for Chemical Desulfurization of Coal: Survey of Thirty-Five Coals. EPA-650/2-74-025-A, U.S. Environmental Protection Agency, Washington, D.C.

93. National Electric Reliability Council, 1980, 1980 Summary of Projected Peak Demand, Generating Capability, and Fossil Fuel Requirements, National Electric Reliability Council, Princeton, NJ.

94. National Coal Association, 1979, Steam-Electric Plant Factors, 1979, National Coal Association, Washington, D.C.

95. G. R. Fryling, 1966, Combustion Engineering, 2nd Ed., Combustion Engineering, Inc., New York, NY.

96. Babcock & Wilcox, 1975, Steam/Its Generation and Use, Babcock & Wilcox Co., New York, NY.

97. G. D. Friedlander, 1980, Sixteenth Steam Station Design Survey, Electrical World, Vol. 194, No. 8, November, pp. 67-82.

98. Department of Energy, 1978, Steam-Electric Plant Construction Cost and Annual Production Expenses 1977, DOE/EIA-0033/3 (77), U.S. Department of Energy, Washington, D.C.; DOE, 1979, Steam-Electric Plant Air and Water Quality Control Data, for the Year Ended December 31, 1976, DOE/FERC 0036, U.S. Department of Energy, Washington, D.C. These are issued annually.

99. Title 40, Code of Federal Regulations, Part 60, Subpart D.

100. Chemical Engineering, 1976, 1977, 1978, 1979, Economic Indicators, Vols. 83, 84, 85, and 86.

101. EPRI, 1978, Technical Assessment Guide, EPRI PS-866-SR, Special Report, Electric Power Research Institute, Palo Alto, CA.

102. E. L. Grant and W. G. Ireson, 1970, Principles of Engineering Economy, Ronald Press, New York, NY.

103. D. O. Seaward, 1981, UOP, Norwalk, Connecticut, Personal Communication.

APPENDIX A

TABLE A-1. ADVANCED LIMESTONE PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	2,552
Feed preparation	4,296
Gas handling	10,476
SO_2 absorption	17,726
Reheat	3,221
Oxidation	2,348
Solids separation	3,456
Total process capital	44,075
Services, utilities, and miscellaneous	2,644
Total direct investment excluding landfill	46,719
Landfill construction	3,362
Landfill equipment	890
Total direct investment	50,971
Indirect Investment	
Engineering design and supervision	3,270
Architect and engineering contractor	934
Construction expense	7,475
Contractor fees	2,336
Contingency	6,073
Waste disposal indirect investment	1,496
Total fixed investment	72,555
Other Capital Investment	
Allowance for startup and modifications	5,345
Interest during construction	11,319
Royalties	0
Land	548
Working capital	3,003
Total capital investment	92,770
Dollars per kW of generation capacity	185.54

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one mile from power plant.

TABLE A-2. ADVANCED LIMESTONE PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	109,100 tons	8.50/ton	927
Adipic acid	138.9 tons	1,500.00/ton	208
Total raw materials cost			1,135
Conversion costs			
Operating labor and supervision			
FGD	42,920 man-hr	15.00/man-hr	644
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	425,200 klb	2.50/klb	1,063
Process water	178,300 kgal	0.14/kgal	25
Electricity	45,024,000 kWh	0.037/kWh	1,666
Diesel fuel	73,150 gal	1.60/gal	117
Maintenance			
Labor and material			3,865
Analysis	3,330 man-hr	21.00/man-hr	70
Total conversion costs			7,974
Total direct costs			9,109
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,062
Total first-year operating and maintenance costs			12,171
Levelized capital charges (14.7% of total capital investment)			13,637
Total first-year annual revenue requirements			25,808
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			22,954
Levelized capital charges (14.7% of total capital investment)			13,637
Levelized annual revenue requirements			36,591

	M$	Mills/kWh
First-year annual revenue requirements	25.81	9.38
Levelized annual revenue requirements	36.59	13.31

Basis: One-year, 5,500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-3. ADVANCED LIMESTONE PROCESS

SUMMARY OF CAPITAL INVESTMENT

(200-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	1,382
Feed preparation	2,788
Gas handling	5,558
SO_2 absorption	9,455
Reheat	1,724
Oxidation	1,351
Solids separation	2,682
Total process capital	24,940
Services, utilities, and miscellaneous	1,497
Total direct investment excluding landfill	26,437
Landfill construction	1,870
Landfill equipment	516
Total direct investment	28,823
Indirect Investment	
Engineering design and supervision	2,115
Architect and engineering contractor	793
Construction expense	4,759
Contractor fees	1,586
Contingency	3,569
Waste disposal indirect investment	881
Total fixed investment	42,526
Other Capital Investment	
Allowance for startup and modifications	3,140
Interest during construction	6,634
Royalties	0
Land	307
Working capital	1,764
Total capital investment	54,371
Dollars per kW of generation capacity	271.85

Basis: North central new power unit with 30-year life. Mid-1982 costs. 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one mile from power plant.

TABLE A-4. ADVANCED LIMESTONE PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	44,500 tons	8.50/ton	379
Adipic acid	56.7 tons	1,500.00/ton	85
Total raw materials cost			464
Conversion costs			
Operating labor and supervision			
FGD	32,840 man-hr	15.00/man-hr	493
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	173,600 klb	2.50/klb	434
Process water	72,800 kgal	0.14/kgal	10
Electricity	19,260,000 kWh	0.037/kWh	713
Diesel fuel	41,890 gal	1.60/gal	67
Maintenance			
Labor and material			2,451
Analysis	3,310 man-hr	21.00/man-hr	69
Total conversion costs			4,761
Total direct costs			5,225
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,122
Total first-year operating and maintenance costs			7,347
Levelized capital charges (14.7% of total capital investment)			7,993
Total first-year annual revenue requirements			15,340
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			13,856
Levelized capital charges (14.7% of total capital investment)			7,993
Levelized annual revenue requirements			21,849

	M$	Mills/kWh
First-year annual revenue requirements	15.34	13.95
Levelized annual revenue requirements	21.85	19.86

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-5. ADVANCED LIMESTONE PROCESS

SUMMARY OF CAPITAL INVESTMENT

(1,000-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	2,917
Feed preparation	5,603
Gas handling	19,863
SO_2 absorption	33,988
Reheat	6,152
Oxidation	4,417
Solids separation	4,382
Total process capital	77,322
Services, utilities, and miscellaneous	4,639
Total direct investment excluding landfill	81,961
Landfill construction	5,221
Landfill equipment	1,466
Total direct investment	88,648
Indirect Investment	
Engineering design and supervision	4,918
Architect and engineering contractor	820
Construction expense	11,475
Contractor fees	3,278
Contingency	10,245
Waste disposal indirect investment	2,214
Total fixed investment	121,598
Other Capital Investment	
Allowance for startup and modifications	9,015
Interest during construction	18,969
Royalties	0
Land	847
Working capital	5,059
Total capital investment	155,488
Dollars per kW of generation capacity	155.49

Basis: North central new power unit with 30-year life. Mid-1982 costs, 2 spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one mile from power plant.

TABLE A-6. ADVANCED LIMESTONE PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1,000-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	211,200 tons	8.50/ton	1,796
Adipic acid	269 tons	1,500.00/ton	403
Total raw materials cost			2,199
Conversion costs			
Operating labor and supervision			
FGD	53,840 man-hr	15.00/man-hr	808
Waste disposal	37,440 man-hr	21.00/man-hr	786
Utilities			
Steam	823,500 klb	2.50/klb	2,059
Process water	345,300 kgal	0.14/kgal	48
Electricity	85,095,000 kWh	0.037/kWh	3,148
Diesel fuel	124,300 gal	1.60/gal	199
Maintenance			
Labor and material			5,938
Analysis	6,660 man-hr	21.00/man-hr	140
Total conversion costs			13,126
Total direct costs			15,325
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,603
Total first-year operating and maintenance costs			19,928
Levelized capital charges (14.7% of total capital investment)			22,857
Total first-year annual revenue requirements			42,785
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			37,584
Levelized capital charges (14.7% of total capital investment)			22,857
Levelized annual revenue requirements			60,441

	M$	Mills/kWh
First-year annual revenue requirements	42.79	7.78
Levelized annual revenue requirements	60.44	10.99

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-7. ADVANCED LIMESTONE PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 2.0% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials handling	1,337
Feed preparation	2,838
Gas handling	9,808
SO_2 absorption	16,607
Reheat	2,593
Oxidation	2,344
Solids separation	2,862
Total process capital	38,389
Services, utilities, and miscellaneous	2,303
Total direct investment excluding landfill	40,692
Landfill construction	2,178
Landfill equipment	576
Total direct investment	43,446
Indirect Investment	
Engineering design and supervision	2,848
Architect and engineering contractor	814
Construction expense	6,511
Contractor fees	2,035
Contingency	5,290
Waste disposal indirect investment	969
Total fixed investment	61,913
Other Capital Investment	
Allowance for startup and modifications	4,654
Interest during construction	9,658
Royalties	0
Land	358
Working capital	2,526
Total capital investment	79,109
Dollars per kW of generation capacity	158.22

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one mile from power plant.

TABLE A-8. ADVANCED LIMESTONE PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 2.0% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	56,900 tons	8.50/ton	484
Adipic acid	71.8 tons	1,500.00/ton	108
Total raw materials cost			592
Conversion costs			
Operating labor and supervision			
FGD	41,590 man-hr	15.00/man-hr	624
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	267,800 klb	2.50/klb	669
Process water	149,100 kgal	1.60/kgal	21
Electricity	35,916,000 kWh	0.037/kWh	1,329
Diesel fuel	47,550 gal	1.60/gal	76
Maintenance			
Labor and material			3,338
Analysis	3,310 man-hr	21.00/man-hr	70
Total conversion costs			6,651
Total direct costs			7,243
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,733
Total first-year operating and maintenance costs			9,976
Levelized capital charges (14.7% of total capital investment)			11,629
Total first-year annual revenue requirements			21,605
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			18,814
Levelized capital charges (14.7% of total capital investment)			11,629
Levelized annual revenue requirements			30,443

	M$	Mills/kWh
First-year annual revenue requirements	21.61	7.86
Levelized annual revenue requirements	30.44	11.07

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-9. ADVANCED LIMESTONE PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 5.0% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	2,630
Feed preparation	4,890
Gas handling	10,510
SO_2 absorption	17,795
Reheat	3,224
Oxidation	2,612
Solids separation	3,922
Total process capital	45,583
Services, utilities, and miscellaneous	2,735
Total direct investment excluding landfill	48,318
Landfill construction	4,274
Landfill equipment	1,267
Total direct investment	53,859
Indirect Investment	
Engineering design and supervision	3,382
Architect and engineering contractor	966
Construction expense	7,731
Contractor fees	2,416
Contingency	6,281
Waste disposal indirect investment	1,929
Total fixed investment	76,564
Other Capital Investment	
Allowance for startup and modifications	5,529
Interest during construction	11,944
Royalties	0
Land	692
Working capital	3,211
Total capital investment	97,940
Dollars per kW of generation capacity	195.88

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one mile from power plant.

TABLE A-10. ADVANCED LIMESTONE PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 5.0% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	156,000 tons	8.50/ton	1,326
Adipic acid	199.3 tons	1,500.00/ton	299
Total raw materials cost			1,625
Conversion costs			
Operating labor and supervision			
FGD	42,990 man-hr	15.00/man-hr	645
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	427,000 klb	2.50/klb	1,068
Process water	192,100 kgal	0.14/kgal	27
Electricity	49,627,000 kWh	0.037/kWh	1,836
Diesel fuel	108,100 gal	1.60/gal	173
Maintenance			
Labor and material			4,032
Analysis	3,350 man-hr	21.00/man-hr	70
Total conversion costs			8,462
Total direct costs			10,087
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,215
Total first-year operating and maintenance costs			13,302
Levelized capital charges (14.7% of total capital investment)			14,397
Total first-year annual revenue requirements			27,699
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			25,088
Levelized capital charges (14.7% of total capital investment)			14,397
Levelized annual revenue requirements			39,485

	M$	Mills/kWh
First-year annual revenue requirements	27.70	10.07
Levelized annual revenue requirements	39.49	14.36

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-11. ADVANCED LIMESTONE PROCESS

SUMMARY OF CAPITAL INVESTMENT

(200-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	1,895
Feed preparation	2,974
Gas handling	6,378
SO_2 absorption	10,636
Reheat	1,940
Oxidation	1,440
Solids separation	2,752
Total process capital	28,015
Services, utilities, and miscellaneous	1,681
Total direct investment excluding landfill	29,696
Landfill construction	1,463
Landfill equipment	518
Total direct investment	31,677
Indirect Investment	
Engineering design and supervision	2,376
Architect and engineering contractor	891
Construction expense	5,345
Contractor fees	1,782
Contingency	4,009
Waste disposal indirect investment	713
Total fixed investment	46,793
Other Capital Investment	
Allowance for startup and modifications	3,528
Interest during construction	7,300
Royalties	0
Land	241
Working capital	1,916
Total capital investment	59,778
Dollars of total capital per kW of generation capacity	298.89

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-12. ADVANCED LIMESTONE PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	45,470 tons	8.50/ton	386
Adipic acid	57.9 tons	1,500.00/ton	87
Total raw materials cost			473
Conversion costs			
Operating labor and supervision			
FGD	32,840 man-hr	15.00/man-hr	493
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	177,200 klb	2.50/klb	443
Process water	74,300 kgal	0.14/kgal	10
Electricity	21,021,000 kWh	0.037/kWh	778
Diesel fuel	41,900 gal	1.60/gal	67
Maintenance			
Labor and material			2,732
Analysis	3,310 man-hr	21.00/man-hr	70
Total conversion costs			5,117
Total direct costs			5,590
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,291
Total first-year operating and maintenance costs			7,881
Levelized capital charges (15.4% of total capital investment)			9,206
Total first-year annual revenue requirements			17,087
Levelized first-year operating and maintenance costs (1.629 first-year O&M)			12,838
Levelized capital charges (15.4% of total capital investment)			9,206
Levelized annual revenue requirements			22,044

	M$	Mills/kWh
First-year annual revenue requirements	17.09	15.53
Levelized annual revenue requirements	22.04	20.04

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-13. ADVANCED LIMESTONE PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	3,362
Feed preparation	5,669
Gas handling	13,814
SO_2 absorption	23,393
Reheat	4,250
Oxidation	3,095
Solids separation	4,546
Total process capital	58,129
Services, utilities, and miscellaneous	3,488
Total direct investment excluding landfill	61,617
Landfill construction	2,612
Landfill equipment	896
Total direct investment	65,125
Indirect Investment	
Engineering design and supervision	4,313
Architect and engineering contractor	1,232
Construction expense	9,859
Contractor fees	3,081
Contingency	8,010
Waste disposal indirect investment	1,203
Total fixed investment	92,823
Other Capital Investment	
Allowance for startup and modifications	7,049
Interest during construction	14,480
Royalties	0
Land	429
Working capital	3,693
Total capital investment	118,474
Dollars of total capital per kW of generation capacity	236.95

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-14. ADVANCED LIMESTONE PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	111,400 tons	8.50/ton	947
Adipic acid	141.8 tons	1,500.00/ton	213
Total raw materials cost			1,160
Conversion costs			
Operating labor and supervision			
FGD	42,920 man-hr	15.00/man-hr	644
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	434,200 klb	2.50/klb	1,086
Process water	182,060 kgal	0.14/kgal	25
Electricity	50,645,000 kWh	0.037/kWh	1,874
Diesel fuel	73,230 gal	1.60/gal	117
Maintenance			
Labor and material			5,035
Analysis	3,330 man-hr	21.00/man-hr	70
Total conversion costs			9,375
Total direct costs			10,535
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,764
Total first-year operating and maintenance costs			14,299
Levelized capital charges (15.4% of total capital investment)			18,245
Total first-year annual revenue requirements			32,544
Levelized first-year operating and maintenance costs (1.629 first-year O&M)			23,293
Levelized capital charges (15.4% of total capital investment)			18,245
Levelized annual revenue requirements			41,538

	M$	Mills/kWh
First-year annual revenue requirements	32.54	11.83
Levelized annual revenue requirements	41.54	15.10

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-15. ADVANCED LIMESTONE PROCESS

SUMMARY OF CAPITAL INVESTMENT

(1,000-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	5,170
Feed preparation	9,199
Gas handling	27,238
SO_2 absorption	46,088
Reheat	8,375
Oxidation	6,105
Solids separation	6,624
Total process capital	108,799
Services, utilities, and miscellaneous	6,528
Total direct investment excluding landfill	115,327
Landfill construction	4,075
Landfill equipment	1,482
Total direct investment	120,884
Indirect Investment	
Engineering design and supervision	6,920
Architect and engineering contractor	1,153
Construction expense	16,146
Contractor fees	4,613
Contingency	14,416
Waste disposal indirect investment	1,795
Total fixed investment	165,927
Other Capital Investment	
Allowance for startup and modifications	12,686
Interest during construction	25,885
Royalties	0
Land	666
Working capital	6,565
Total capital investment	211,729
Dollars of total capital per kW of generation capacity	211.73

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-16. ADVANCED LIMESTONE PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1,000-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	218,200 tons	8.50/ton	1,855
Adipic acid	277.8 tons	1,500.00/ton	417
Total raw materials cost			2,272
Conversion costs			
Operating labor and supervision			
FGD	53,850 man-hr	15.00/man-hr	808
Solids disposal	37,740 man-hr	21.00/man-hr	793
Utilities			
Steam	850,400 klb	2.50/klb	2,126
Process water	356,600 kgal	0.14/kgal	50
Electricity	98,613,000 kWh	0.037/kWh	3,649
Diesel fuel	125,030 gal	1.60/gal	200
Maintenance			
Labor and material			8,240
Analysis	6,660 man-hr	21.00/man-hr	140
Total conversion costs			16,006
Total direct costs			18,278
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			5,989
Total first-year operating and maintenance costs			24,267
Levelized capital charges (15.4% of total capital investment)			32,606
Total first-year annual revenue requirements			56,873
Levelized first-year operating and maintenance costs (1.629 first-year O&M)			39,531
Levelized capital charges (15.4% of total capital investment)			32,606
Levelized annual revenue requirements			72,137

	M$	Mills/kWh
First-year annual revenue requirements	56.87	10.34
Levelized annual revenue requirements	72.14	13.12

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-17. DAVY S-H PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	3,476
Feed preparation	1,242
Gas handling	18,616
SO_2 absorption	14,665
Reheat	4,157
Oxidation	5,039
Solids separation	2,565
Total process capital	49,760
Services, utilities, and miscellaneous	2,986
Total direct investment excluding landfill	52,746
Landfill construction	3,279
Landfill equipment	880
Total direct investment	56,905
Indirect Investment	
Engineering design and supervision	3,692
Architect and engineering contractor	1,055
Construction expense	8,439
Contractor fees	2,637
Contingency	13,714
Waste disposal indirect investment	1,462
Total fixed investment	87,904
Other Capital Investment	
Allowance for startup and modifications	8,228
Interest during construction	13,713
Royalties	1,660
Land	544
Working capital	3,477
Total capital investment	115,526
Dollars of total capital per kW of generation capacity	231.05

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-18. DAVY S-H PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	57,970 tons	75.00/ton	4,348
Formic acid	222.8 tons	760/ton	169
Ferric chloride	132 anhyd. tons	234/anhyd. ton	31
Antifoulant	0.83 tons	2,840/ton	2
Total raw materials cost			4,550
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	572,000 klb	2.50/klb	1,430
Process water	120,800 kgal	0.14/kgal	17
Electricity	60,540,000 kWh	0.037/kWh	2,240
Diesel fuel	72,790 gal	1.60/gal	116
Maintenance			
Labor and material			3,290
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			8,380
Total direct costs			12,930
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,746
Total first-year operating and maintenance costs			15,676
Levelized capital charges (14.7% of total capital investment)			16,982
Total first-year annual revenue requirements			32,658
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			29,565
Levelized capital charges (14.7% of total capital investment)			16,982
Levelized annual revenue requirements			46,547

	M$	Mills/kWh
First-year annual revenue requirements	32.7	11.88
Levelized annual revenue requirements	46.5	16.93

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-19. DAVY S-H PROCESS

SUMMARY OF CAPITAL INVESTMENT

(200-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	1,908
Feed preparation	719
Gas handling	9,734
SO_2 absorption	7,275
Reheat	2,160
Oxidation	2,635
Solids separation	1,654
Total process capital	26,085
Services, utilities, and miscellaneous	1,565
Total direct investment excluding landfill	27,650
Landfill construction	1,826
Landfill equipment	512
Total direct investment	29,988
Indirect Investment	
Engineering design and supervision	2,212
Architect and engineering contractor	830
Construction expense	4,977
Contractor fees	1,659
Contingency	7,466
Waste disposal indirect investment	862
Total fixed investment	47,994
Other Capital Investment	
Allowance for startup and modifications	4,479
Interest during construction	7,487
Royalties	821
Land	297
Working capital	1,823
Total capital investment	62,951
Dollars of total capital per kW of generation capacity	314.76

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-20. DAVY S-H PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	23,670 tons	75.00/ton	1,776
Formic acid	90.9 tons	760/ton	69
Ferric chloride	54 anhyd. tons	234/anhyd. ton	13
Antifoulant	0.33 tons	2,840/ton	1
Total raw materials cost			1,859
Conversion costs			
Operating labor and supervision			
FGD	33,470 man-hr	15.00/man-hr	502
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	233,600 klb	2.50/klb	584
Process water	50,000 kgal	0.14/kgal	7
Electricity	25,076,000 kWh	0.037/kWh	928
Diesel fuel	41,880 gal	1.60/gal	67
Maintenance			
Labor and material			2,006
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			4,687
Total direct costs			6,546
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,861
Total first-year operating and maintenance costs			8,407
Levelized capital charges (14.7% of total capital investment)			9,254
Total first-year annual revenue requirements			17,661
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			15,856
Levelized capital charges (14.7% of total capital investment)			9,254
Levelized annual revenue requirements			25,110

	M$	Mills/kWh
First-year annual revenue requirements	17.7	16.06
Levelized annual revenue requirements	25.1	22.83

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-21. DAVY S-H PROCESS

SUMMARY OF CAPITAL INVESTMENT

(1,000-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	5,412
Feed preparation	1,859
Gas handling	36,428
SO_2 absorption	28,457
Reheat	8,126
Oxidation	9,860
Solids separation	3,546
Total process capital	93,688
Services, utilities, and miscellaneous	5,621
Total direct investment excluding landfill	99,309
Landfill construction	5,091
Landfill equipment	1,454
Total direct investment	105,854
Indirect Investment	
Engineering design and supervision	5,959
Architect and engineering contractor	993
Construction expense	13,903
Contractor fees	3,972
Contingency	24,827
Waste disposal indirect investment	2,164
Total fixed investment	157,672
Other Capital Investment	
Allowance for startup and modifications	14,896
Interest during construction	24,597
Royalties	2,444
Land	858
Working capital	6,191
Total capital investment	206,658
Dollars of total capital per kW of generation capacity	206.66

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-22. DAVY S-H PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1,000-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	112,280 tons	75.00/ton	8,421
Formic acid	431.5 tons	760/ton	328
Ferric chloride	256 anhyd. tons	234/anhyd. ton	60
Antifoulant	1.6 tons	2,840/ton	5
Total raw materials cost			8,814
Conversion costs			
Operating labor and supervision			
FGD	55,130 man-hr	15.00/man-hr	827
Solids disposal	33,280 man-hr	21.00/man-hr	699
Utilities			
Steam	1,108,000 klb	2.50/klb	2,770
Process water	238,500 kgal	0.14/kgal	33
Electricity	116,680,000 kWh	0.037/kWh	4,322
Diesel fuel	123,100 gal	1.60/gal	197
Maintenance			
Labor and material			5,162
Analysis	6,630 man-hr	21.00/man-hr	140
Total conversion costs			14,150
Total direct costs			22,964
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,097
Total first-year operating and maintenance costs			27,061
Levelized capital charges (14.7% of total capital investment)			30,379
Total first-year annual revenue requirements			57,440
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			51,037
Levelized capital charges (14.7% of total capital investment)			30,379
Levelized annual revenue requirements			81,416

	M$	Mills/kWh
First-year annual revenue requirements	57.4	10.44
Levelized annual revenue requirements	81.4	14.80

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-23. DAVY S-H PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 2.0% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	2,231
Feed preparation	830
Gas handling	17,218
SO_2 absorption	13,146
Reheat	3,831
Oxidation	4,660
Solids separation	1,855
Total process capital	43,771
Services, utilities, and miscellaneous	2,626
Total direct investment excluding landfill	46,397
Landfill construction	2,124
Landfill equipment	571
Total direct investment	49,092
Indirect Investment	
Engineering design and supervision	3,248
Architect and engineering contractor	928
Construction expense	7,424
Contractor fees	2,320
Contingency	12,063
Waste disposal indirect investment	946
Total fixed investment	76,021
Other Capital Investment	
Allowance for startup and modifications	7,238
Interest during construction	11,859
Royalties	1,493
Land	361
Working capital	2,871
Total capital investment	99,843
Dollars of total capital per kW of generation capacity	199.69

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-24. DAVY S-H PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 2.0% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	29,920 tons	75.00/ton	2,244
Formic acid	115.0 tons	760/ton	87
Ferric chloride	68 anhyd. tons	234/anhyd. ton	16
Antifoulant	0.43 tons	2,840/ton	1
Total raw materials cost			2,348
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	369,400 klb	2.50/klb	924
Process water	108,700 kgal	0.14/kgal	15
Electricity	53,220,000 kWh	0.037/kWh	1,969
Diesel fuel	47,400 gal	1.60/gal	76
Maintenance			
Labor and material			2,865
Analysis	4,960 man-hr	21.00/man-hr	106
Total conversion costs			7,135
Total direct costs			9,483
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,491
Total first-year operating and maintenance costs			11,974
Levelized capital charges (14.7% of total capital investment)			14,677
Total first-year annual revenue requirements			26,651
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			22,583
Levelized capital charges (14.7% of total capital investment)			14,677
Levelized annual revenue requirements			37,260

	M$	Mills/kWh
First-year annual revenue requirements	26.7	9.69
Levelized annual revenue requirements	37.3	13.55

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-25. DAVY S-H PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 5.0% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	4,427
Feed preparation	1,548
Gas handling	18,572
SO_2 absorption	14,616
Reheat	4,147
Oxidation	5,027
Solids separation	3,061
Total process capital	51,398
Services, utilities, and miscellaneous	3,084
Total direct investment excluding landfill	54,482
Landfill construction	4,168
Landfill equipment	1,179
Total direct investment	59,829
Indirect Investment	
Engineering design and supervision	3,814
Architect and engineering contractor	1,090
Construction expense	8,717
Contractor fees	2,724
Contingency	14,165
Waste disposal indirect investment	1,869
Total fixed investment	92,208
Other Capital Investment	
Allowance for startup and modifications	8,499
Interest during construction	14,384
Royalties	1,847
Land	684
Working capital	3,785
Total capital investment	121,407
Dollars of total capital per kW of generation capacity	242.81

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-26. DAVY S-H PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 5.0% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	83,190 tons	75.00/ton	6,239
Formic acid	319.7 tons	760/ton	243
Ferric chloride	187 anhyd. tons	234/anhyd. ton	44
Antifoulant	1.2 tons	2,840/ton	3
Total raw materials cost			6,529
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	29,120 man-hr	21.00/man-hr	612
Utilities			
Steam	570,000 klb	2.50/klb	1,425
Process water	124,000 kgal	0.14/kgal	17
Electricity	61,404,000 kWh	0.037/kWh	2,272
Diesel fuel	105,200 gal	1.60/gal	168
Maintenance			
Labor and material			3,429
Analysis	5,000 man-hr	21.00/man-hr	105
Total conversion costs			8,686
Total direct costs			15,215
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,882
Total first-year operating and maintenance costs			18,097
Levelized capital charges (14.7% of total capital investment)			17,847
Total first-year annual revenue requirements			35,944
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			34,131
Levelized capital charges (14.7% of total capital investment)			17,847
Levelized annual revenue requirements			51,978

	M$	Mills/kWh
First-year annual revenue requirements	35.9	13.07
Levelized annual revenue requirements	52.0	18.90

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-27. DAVY S-H PROCESS

SUMMARY OF CAPITAL INVESTMENT

(200-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	2,514
Feed preparation	947
Gas handling	12,830
SO_2 absorption	9,640
Reheat	2,849
Oxidation	3,473
Solids separation	2,172
Total process capital	34,425
Services, utilities, and miscellaneous	2,066
Total direct investment excluding landfill	36,491
Landfill construction	1,427
Landfill equipment	514
Total direct investment	38,432
Indirect Investment	
Engineering design and supervision	2,919
Architect and engineering contractor	1,095
Construction expense	6,568
Contractor fees	2,189
Contingency	9,852
Waste disposal indirect investment	697
Total fixed investment	61,752
Other Capital Investment	
Allowance for startup and modifications	5,911
Interest during construction	9,633
Royalties	821
Land	234
Working capital	2,255
Total capital investment	80,606
Dollars of total capital per kW of generation capacity	403.03

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-28. DAVY S-H PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	24,160 tons	75.00/ton	1,812
Formic acid	92.9 tons	760/ton	71
Ferric chloride	55 anhyd. tons	234/anhyd. ton	13
Antifoulant	0.35 tons	2,840/ton	1
Total raw materials cost			1,897
Conversion costs			
Operating labor and supervision			
FGD	33,470 man-hr	15.00/man-hr	502
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	238,400 klb	2.50/klb	596
Process water	51,300 kgal	0.14/kgal	8
Electricity	25,596,000 kWh	0.037/kWh	947
Diesel fuel	41,790 gal	1.60/gal	66
Maintenance			
Labor and material			2,613
Analysis	3,330 man-hr	21.00/man-hr	70
Total conversion costs			5,326
Total direct costs			7,223
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,225
Total first-year operating and maintenance costs			9,448
Levelized capital charges (14.7% of total capital investment)			12,413
Total first-year annual revenue requirements			21,861
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			15,391
Levelized capital charges (14.7% of total capital investment)			12,413
Levelized annual revenue requirements			27,804

	M$	Mills/kWh
First-year annual revenue requirements	21.9	19.87
Levelized annual revenue requirements	27.8	25.28

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-29. DAVY S-H PROCESS
SUMMARY OF CAPITAL INVESTMENT
(500-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	4,582
Feed preparation	1,635
Gas handling	24,547
SO_2 absorption	19,442
Reheat	5,485
Oxidation	6,644
Solids separation	3,369
Total process capital	65,704
Services, utilities, and miscellaneous	3,942
Total direct investment excluding landfill	69,646
Landfill construction	2,545
Landfill equipment	885
Total direct investment	73,076
Indirect Investment	
Engineering design and supervision	4,875
Architect and engineering contractor	1,393
Construction expense	11,143
Contractor fees	3,482
Contingency	18,108
Waste disposal indirect investment	1,174
Total fixed investment	113,251
Other Capital Investment	
Allowance for startup and modifications	10,865
Interest during construction	17,667
Royalties	1,660
Land	428
Working capital	4,179
Total capital investment	148,050
Dollars of total capital per kW of generation capacity	296.01

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-30. DAVY S-H PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	59,190 tons	75.00/ton	4,439
Formic acid	227.5 tons	760/ton	173
Ferric chloride	135 anhyd. tons	234/anhyd. ton	32
Antifoulant	0.85 tons	2,840/ton	2
Total raw materials cost			4,646
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	584,000 klb	2.50/klb	1,460
Process water	126,000 kgal	0.14/kgal	18
Electricity	61,852,000 kWh	0.037/kWh	2,289
Diesel fuel	72,900 gal	1.60/gal	117
Maintenance			
Labor and material			4,282
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			9,453
Total direct costs			14,099
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,341
Total first-year operating and maintenance costs			17,440
Levelized capital charges (14.7% of total capital investment)			22,800
Total first-year annual revenue requirements			40,240
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			28,410
Levelized capital charges (14.7% of total capital investment)			22,800
Levelized annual revenue requirements			51,210

	M$	Mills/kWh
First-year annual revenue requirements	40.2	14.63
Levelized annual revenue requirements	51.2	18.62

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-31. DAVY S-H PROCESS

SUMMARY OF CAPITAL INVESTMENT

(1,000-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	7,189
Feed preparation	2,464
Gas handling	48,401
SO_2 absorption	38,128
Reheat	10,808
Oxidation	13,100
Solids separation	4,683
Total process capital	124,773
Services, utilities, and miscellaneous	7,486
Total direct investment excluding landfill	132,259
Landfill construction	3,969
Landfill equipment	1,463
Total direct investment	137,691
Indirect Investment	
Engineering design and supervision	7,936
Architect and engineering contractor	1,323
Construction expense	18,516
Contractor fees	5,290
Contingency	33,065
Waste disposal indirect investment	1,754
Total fixed investment	205,575
Other Capital Investment	
Allowance for startup and modifications	19,839
Interest during construction	32,070
Royalties	2,444
Land	682
Working capital	7,539
Total capital investment	268,149
Dollars of total capital per kW of generation capacity	268.15

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-32. DAVY S-H PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1,000-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	115,940 tons	75.00/ton	8,695
Formic acid	445.5 tons	760/ton	339
Ferric chloride	264 anhyd. tons	234/anhyd. ton	62
Antifoulant	1.65 tons	2,840/ton	5
Total raw materials cost			9,101
Conversion costs			
Operating labor and supervision			
FGD	55,140 man-hr	15.00/man-hr	827
Solids disposal	37,440 man-hr	21.00/man-hr	786
Utilities			
Steam	1,144,000 klb	2.50/klb	2,860
Process water	246,700 kgal	0.14/kgal	35
Electricity	120,570,000 kWh	0.037/kWh	4,461
Diesel fuel	123,800 gal	1.60/gal	198
Maintenance			
Labor and material			6,776
Analysis	6,660 man-hr	21.00/man-hr	140
Total conversion costs			16,083
Total direct costs			25,184
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			5,117
Total first-year operating and maintenance costs			30,301
Levelized capital charges (14.7% of total capital investment)			41,295
Total first-year annual revenue requirements			71,596
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			49,360
Levelized capital charges (14.7% of total capital investment)			41,295
Levelized annual revenue requirements			90,655

	M$	Mills/kWh
First-year annual revenue requirements	71.6	13.02
Levelized annual revenue requirements	90.6	16.48

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-33. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	2,915
Feed preparation	5,307
Gas handling	14,296
SO_2 absorption	17,550
Reheat	3,950
Oxidation	4,748
Neutralization	557
Chloride purge	834
Solids separation	2,733
Total process capital	52,890
Services, utilities, and miscellaneous	3,173
Total direct investment excluding landfill	56,063
Landfill construction	3,307
Landfill equipment	883
Total direct investment	60,253
Indirect Investment	
Engineering design and supervision	3,924
Architect and engineering contractor	1,121
Construction expense	8,970
Contractor fees	3,364
Contingency	14,689
Waste disposal indirect investment	1,473
Total fixed investment	93,794
Other Capital Investment	
Allowance for startup and modifications	8,813
Interest during construction	14,632
Royalties	0
Land	542
Working capital	3,554
Total capital investment	121,335
Dollars of total capital per kW of generation capacity	242.67

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-34. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	107,200 tons	8.50/ton	911
Aluminum sulfate	1,392 tons	350.00/ton	487
Total raw materials cost			1,398
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	576,900 klb	2.50/klb	1,442
Process water	210,000 kgal	0.14/kgal	29
Electricity	53,836,000 kWh	0.037/kWh	1,992
Diesel fuel	72,900 gal	1.60/gal	117
Maintenance			
Labor and material			4,611
Analysis	5,720 man-hr	21.00/man-hr	120
Total conversion costs			9,493
Total direct costs			10,891
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,548
Total first-year operating and maintenance costs			14,439
Levelized capital charges (14.7% of total capital investment)			17,836
Total first-year annual revenue requirements			32,275
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			27,232
Levelized capital charges (14.7% of total capital investment)			17,836
Levelized annual revenue requirements			45,068

	M$	Mills/kWh
First-year annual revenue requirements	32.3	11.74
Levelized annual revenue requirements	45.1	16.39

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-35. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(200-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	1,348
Feed preparation	3,003
Gas handling	7,475
SO_2 absorption	9,029
Reheat	2,048
Oxidation	2,514
Neutralization	298
Chloride purge	487
Solids separation	1,611
Total process capital	27,813
Services, utilities, and miscellaneous	1,669
Total direct investment excluding landfill	29,482
Landfill construction	1,840
Landfill equipment	514
Total direct investment	31,836
Indirect Investment	
Engineering design and supervision	2,359
Architect and engineering contractor	884
Construction expense	5,307
Contractor fees	2,064
Contingency	8,019
Waste disposal indirect investment	868
Total fixed investment	51,337
Other Capital Investment	
Allowance for startup and modifications	4,812
Interest during construction	8,009
Royalties	0
Land	296
Working capital	1,957
Total capital investment	66,411
Dollars of total capital per kW of generation capacity	332.06

Basis: North central new power unit with 30-year life. Mid-1982 costs. 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-36. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	43,750 tons	8.50/ton	372
Aluminum sulfate	568 tons	350.00/ton	199
Total raw materials cost			571
Conversion costs			
Operating labor and supervision			
FGD	33,470 man-hr	15.00/man-hr	502
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	235,600 klb	2.50/klb	589
Process water	85,700 kgal	0.14/kgal	12
Electricity	22,562,000 kWh	0.037/kWh	835
Diesel fuel	41,820 gal	1.60/gal	67
Maintenance			
Labor and material			2,724
Analysis	3,810 man-hr	21.00/man-hr	80
Total conversion costs			5,333
Total direct costs			5,904
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,298
Total first-year operating and maintenance costs			8,202
Levelized capital charges (14.7% of total capital investment)			9,762
Total first-year annual revenue requirements			17,964
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			15,469
Levelized capital charges (14.7% of total capital investment)			9,762
Levelized annual revenue requirements			25,231

	M$	Mills/kWh
First-year annual revenue requirements	18.0	16.33
Levelized annual revenue requirements	25.2	22.94

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-37. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(1,000-MW, new power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	3,944
Feed preparation	9,210
Gas handling	27,974
SO_2 absorption	34,253
Reheat	7,719
Oxidation	9,309
Neutralization	1,093
Chloride purge	1,240
Solids separation	4,036
Total process capital	98,778
Services, utilities, and miscellaneous	5,927
Total direct investment excluding landfill	104,705
Landfill construction	5,134
Landfill equipment	1,462
Total direct investment	111,301
Indirect Investment	
Engineering design and supervision	6,282
Architect and engineering contractor	1,047
Construction expense	14,659
Contractor fees	5,235
Contingency	26,386
Waste disposal indirect investment	2,181
Total fixed investment	167,091
Other Capital Investment	
Allowance for startup and modifications	15,831
Interest during construction	26,066
Royalties	0
Land	853
Working capital	6,273
Total capital investment	216,114
Dollars of total capital per kW of generation capacity	216.11

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-38. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1,000-MW, new power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	207,500 tons	8.50/ton	1,763
Aluminum sulfate	2,695 tons	350.00/ton	943
Total raw materials cost			2,706
Conversion costs			
Operating labor and supervision			
FGD	55,140 man-hr	15.00/man-hr	827
Solids disposal	33,280 man-hr	21.00/man-hr	699
Utilities			
Steam	1,117,000 klb	2.50/klb	2,794
Process water	406,500 kgal	0.14/kgal	57
Electricity	104,760,000 kWh	0.037/kWh	3,876
Diesel fuel	123,500 gal	1.60/gal	198
Maintenance			
Labor and material			7,527
Analysis	7,650 man-hr	21.00/man-hr	161
Total conversion costs			16,139
Total direct costs			18,845
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			5,528
Total first-year operating and maintenance costs			24,373
Levelized capital charges (14.7% of total capital investment)			31,769
Total first-year annual revenue requirements			56,142
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			45,967
Levelized capital charges (14.7% of total capital investment)			31,769
Levelized annual revenue requirements			77,736

	M$	Mills/kWh
First-year annual revenue requirements	56.1	10.21
Levelized annual revenue requirements	77.7	14.13

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-39. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 2.0% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	1,583
Feed preparation	3,554
Gas handling	13,222
SO_2 absorption	16,083
Reheat	3,637
Oxidation	4,422
Neutralization	522
Chloride purge	561
Solids separation	1,849
Total process capital	45,433
Services, utilities, and miscellaneous	2,726
Total direct investment excluding landfill	48,159
Landfill construction	2,141
Landfill equipment	573
Total direct investment	50,873
Indirect Investment	
Engineering design and supervision	3,371
Architect and engineering contractor	963
Construction expense	7,705
Contractor fees	2,890
Contingency	12,618
Waste disposal indirect investment	953
Total fixed investment	79,373
Other Capital Investment	
Allowance for startup and modifications	7,571
Interest during construction	12,382
Royalties	0
Land	358
Working capital	2,964
Total capital investment	102,648
Dollars of total capital per kW of generation capacity	205.30

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-40. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 2.0% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	55,270 tons	8.50/ton	470
Aluminum sulfate	718 tons	350.00/ton	251
Total raw materials cost			721
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	372,500 klb	2.50/klb	931
Process water	187,100 kgal	0.14/kgal	26
Electricity	43,456,000 kWh	0.037/kWh	1,608
Diesel fuel	47,450 gal	1.60/gal	76
Maintenance			
Labor and material			3,934
Analysis	5,700 man-hr	21.00/man-hr	120
Total conversion costs			7,877
Total direct costs			8,598
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,142
Total first-year operating and maintenance costs			11,740
Levelized capital charges (14.7% of total capital investment)			15,089
Total first-year annual revenue requirements			26,829
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			22,142
Levelized capital charges (14.7% of total capital investment)			15,089
Levelized annual revenue requirements			37,231

	M$	Mills/kWh
First-year annual revenue requirements	26.8	9.76
Levelized annual revenue requirements	37.2	13.54

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-41. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 5.0% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	3,207
Feed preparation	7,422
Gas handling	14,262
SO_2 absorption	17,503
Reheat	3,940
Oxidation	4,738
Neutralization	556
Chloride purge	1,036
Solids separation	3,382
Total process capital	56,046
Services, utilities, and miscellaneous	3,363
Total direct investment excluding landfill	59,409
Landfill construction	4,206
Landfill equipment	1,255
Total direct investment	64,870
Indirect Investment	
Engineering design and supervision	4,159
Architect and engineering contractor	1,188
Construction expense	9,505
Contractor fees	3,565
Contingency	15,565
Waste disposal indirect investment	1,898
Total fixed investment	100,750
Other Capital Investment	
Allowance for startup and modifications	9,339
Interest during construction	15,717
Royalties	0
Land	684
Working capital	3,815
Total capital investment	130,305
Dollars of total capital per kW of generation capacity	260.61

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-42. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 5.0% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	153,800 tons	8.50/ton	1,307
Aluminum sulfate	1,997 tons	350.00/ton	699
Total raw materials cost			2,006
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	29,120 man-hr	21.00/man-hr	612
Utilities			
Steam	574,900 klb	2.50/klb	1,437
Process water	210,600 kgal	0.14/kgal	29
Electricity	59,534,000 kWh	0.037/kWh	2,203
Diesel fuel	107,700 gal	1.60/gal	172
Maintenance			
Labor and material			4,717
Analysis	5,740 man-hr	21.00/man-hr	121
Total conversion costs			9,949
Total direct costs			11,955
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,665
Total first-year operating and maintenance costs			15,620
Levelized capital charges (14.7% of total capital investment)			19,155
Total first-year annual revenue requirements			34,755
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			29,459
Levelized capital charges (14.7% of total capital investment)			19,155
Levelized annual revenue requirements			48,614

	M$	Mills/kWh
First-year annual revenue requirements	34.8	12.65
Levelized annual revenue requirements	48.6	17.68

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-43. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(200-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	1,777
Feed preparation	3,962
Gas handling	9,853
SO_2 absorption	11,920
Reheat	2,702
Oxidation	3,309
Neutralization	392
Chloride purge	641
Solids separation	2,120
Total process capital	36,676
Services, utilities, and miscellaneous	2,201
Total direct investment excluding landfill	38,877
Landfill construction	1,439
Landfill equipment	516
Total direct investment	40,832
Indirect Investment	
Engineering design and supervision	3,110
Architect and engineering contractor	1,166
Construction expense	6,998
Contractor fees	2,721
Contingency	10,575
Waste disposal indirect investment	701
Total fixed investment	66,103
Other Capital Investment	
Allowance for startup and modifications	6,345
Interest during construction	10,312
Royalties	0
Land	232
Working capital	2,398
Total capital investment	85,390
Dollars of total capital per kW of generation capacity	426.95

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-44. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	44,700 tons	8.50/ton	380
Aluminum sulfate	580 tons	350.00/ton	203
Total raw materials cost			583
Conversion costs			
Operating labor and supervision			
FGD	33,470 man-hr	15.00/man-hr	502
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	240,500 klb	2.50/klb	601
Process water	87,500 kgal	0.14/kgal	12
Electricity	23,005,000 kWh	0.037/kWh	851
Diesel fuel	41,820 gal	1.60/gal	67
Maintenance			
Labor and material			3,558
Analysis	3,810 man-hr	21.00/man-hr	80
Total conversion costs			6,195
Total direct costs			6,778
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,798
Total first-year operating and maintenance costs			9,576
Levelized capital charges (15.4% of total capital investment)			13,150
Total first-year annual revenue requirements			22,726
Levelized first-year operating and maintenance costs (1.629 first-year O&M)			15,599
Levelized capital charges (15.4% of total capital investment)			13,150
Levelized annual revenue requirements			28,748

	M$	Mills/kWh
First-year annual revenue requirements	22.7	20.66
Levelized annual revenue requirements	28.7	26.14

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-45. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	3,297
Feed preparation	7,552
Gas handling	18,850
SO_2 absorption	23,179
Reheat	5,212
Oxidation	6,253
Neutralization	733
Chloride purge	1,098
Solids separation	3,596
Total process capital	69,770
Services, utilities, and miscellaneous	4,186
Total direct investment excluding landfill	73,956
Landfill construction	2,566
Landfill equipment	889
Total direct investment	77,411
Indirect Investment	
Engineering design and supervision	5,177
Architect and engineering contractor	1,479
Construction expense	11,833
Contractor fees	4,437
Contingency	19,376
Waste disposal indirect investment	1,183
Total fixed investment	120,896
Other Capital Investment	
Allowance for startup and modifications	11,626
Interest during construction	18,860
Royalties	0
Land	426
Working capital	4,365
Total capital investment	156,173
Dollars of total capital per kW of generation capacity	312.35

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-46. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	109,400 tons	8.50/ton	930
Aluminum sulfate	1,421 tons	350.00/ton	497
Total raw materials cost			1,427
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	589,100 klb	2.50/klb	1,473
Process water	214,300 kgal	0.14/kgal	30
Electricity	55,505,000 kWh	0.037/kWh	2,054
Diesel fuel	72,980 gal	1.60/gal	117
Maintenance			
Labor and material			6,020
Analysis	5,720 man-hr	21.00/man-hr	120
Total conversion costs			10,996
Total direct costs			12,423
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,393
Total first-year operating and maintenance costs			16,816
Levelized capital charges (14.7% of total capital investment)			24,051
Total first-year annual revenue requirements			40,867
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			27,393
Levelized capital charges (14.7% of total capital investment)			24,051
Levelized annual revenue requirements			51,444

	M$	Mills/kWh
First-year annual revenue requirements	40.9	14.86
Levelized annual revenue requirements	51.4	18.71

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-47. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(1,000-MW, existing power unit, 3.5% sulfur in coal)

	Capital investment, k$
Direct Investment	
Materials Handling	5,243
Feed preparation	12,253
Gas handling	37,169
SO_2 absorption	45,629
Reheat	10,270
Oxidation	12,345
Neutralization	1,448
Chloride purge	1,644
Solids separation	5,347
Total process capital	131,348
Services, utilities, and miscellaneous	7,881
Total direct investment excluding landfill	139,229
Landfill construction	4,002
Landfill equipment	1,469
Total direct investment	144,700
Indirect Investment	
Engineering design and supervision	8,354
Architect and engineering contractor	1,392
Construction expense	19,492
Contractor fees	6,961
Contingency	35,086
Waste disposal indirect investment	1,766
Total fixed investment	217,751
Other Capital Investment	
Allowance for startup and modifications	21,051
Interest during construction	33,969
Royalties	0
Land	675
Working capital	7,804
Total capital investment	281,250
Dollars of total capital per kW of generation capacity	281.25

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-48. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1,000-MW, existing power unit, 3.5% sulfur in coal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	214,200 tons	8.50/ton	1,821
Aluminum sulfate	2,783 tons	350.00/ton	974
Total raw materials cost			2,795
Conversion costs			
Operating labor and supervision			
FGD	55,140 man-hr	15.00/man-hr	827
Solids disposal	37,440 man-hr	21.00/man-hr	786
Utilities			
Steam	1,154,000 klb	2.50/klb	2,885
Process water	420,000 kgal	0.14/kgal	59
Electricity	108,133,000 kWh	0.037/kWh	4,001
Diesel fuel	124,200 gal	1.60/gal	199
Maintenance			
Labor and material			9,910
Analysis	7,660 man-hr	21.00/man-hr	161
Total conversion costs			18,828
Total direct costs			21,623
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			7,010
Total first-year operating and maintenance costs			28,633
Levelized capital charges (14.7% of total capital investment)			43,313
Total first-year annual revenue requirements			71,946
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			46,643
Levelized capital charges (14.7% of total capital investment)			43,313
Levelized annual revenue requirements			89,956

	M$	Mills/kWh
First-year annual revenue requirements	71.9	13.08
Levelized annual revenue requirements	90.0	16.36

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE A-49. DOWA PROCESS

SUMMARY OF CAPITAL INVESTMENT

(500-MW, new power unit, 3.5% sulfur in coal, L/G ratio = 80 gal/kft^3)

	Capital investment, k$
Direct Investment	
Materials Handling	2,915
Feed preparation	5,307
Gas handling	14,296
SO_2 absorption	19,799
Reheat	3,950
Oxidation	5,233
Neutralization	557
Chloride purge	834
Solids separation	2,733
Total process capital	55,624
Services, utilities, and miscellaneous	3,337
Total direct investment excluding landfill	58,961
Landfill construction	3,307
Landfill equipment	883
Total direct investment	63,151
Indirect Investment	
Engineering design and supervision	4,127
Architect and engineering contractor	1,179
Construction expense	9,434
Contractor fees	2,948
Contingency	15,330
Waste disposal indirect investment	1,473
Total fixed investment	97,642
Other Capital Investment	
Allowance for startup and modifications	9,198
Interest during construction	15,232
Royalties	0
Land	543
Working capital	3,734
Total capital investment	126,349
Dollars of total capital per kW of generation capacity	252.70

Basis: North central new power unit with 30-year life. Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas. Reheat to 175°F as required. Capital costs begin with an inlet plenum downstream of the ESP. Stack plenum and stack excluded. Disposal site one-mile from power plant.

TABLE A-50. DOWA PROCESS

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, new power unit, 3.5% sulfur in coal, L/G ratio = 80 gal/kft^3)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	107,200 tons	8.50/ton	911
Aluminum sulfate	1,392 tons	350.00/ton	487
Total raw materials cost			1,398
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	576,900 klb	2.50/klb	1,442
Process water	210,000 kgal	0.14/kgal	29
Electricity	63,925,000 kWh	0.037/kWh	2,365
Diesel fuel	72,900 gal	1.60/gal	117
Maintenance			
Labor and material			4,843
Analysis	5,720 man-hr	21.00/man-hr	120
Total conversion costs			10,098
Total direct costs			11,496
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,687
Total first-year operating and maintenance costs			15,183
Levelized capital charges (14.7% of total capital investment)			18,573
Total first-year annual revenue requirements			33,756
Levelized first-year operating and maintenance costs (1.886 first-year O&M)			28,635
Levelized capital charges (14.7% of total capital investment)			18,573
Levelized annual revenue requirements			47,208

	M$	Mills/kWh
First-year annual revenue requirements	33.8	12.27
Levelized annual revenue requirements	47.2	17.17

Basis: One-year, 5500-hour full-load operation of system as described in capital investment table. Mid-1984 cost basis.